Carnation Foods Ltd.
Energy, Mines, and Resources Canada
Manitoba Energy and Mines

Ethanol from Potato Wastes

FEBRUARY, 1988

WARDROP ENGINEERING INC.

SUMMARY

Production of fuel or food/industrial grade ethanol from potato wastes at Carnation Foods' Carberry, Manitoba Plant offers excellent economic potential. It is estimated that seven million litres of anhydrous ethanol can be produced annually, although production could be as high as nine millions litres per year. This would depend on the waste streams selected and on supplementation of the feedstock with waste starch from a Winnipeg potato processing plant. Price of $0.30/L for fuel ethanol, and $0.35/L for food/industrial ethanol could yield an after-tax internal rates of return of 25-35 percent.

Recovery of fermentable sugars from dilute waste streams through reverse osmosis has been determined to be desirable and consistent with the ethanol plant objectives. It will result in an increased capacity, will make acceptable quality water available for recycle, and will help offset the organic loading in the ethanol plant effluents. If these dilute wastes are neglected, production would drop by up to 800,000 L of ethanol per year, and a price increase of $0.05/L would be required to maintain an after-tax IRR of 25-30 percent.

The project viability is enhanced if feedstock is supplemented with starch from a Winnipeg potato processing plant. This starch appears to be available at the cost of shipping it from Winnipeg to Carberry. Use of vapour recompression and heat pumps to replace steam for distillation has not been found to be economically viable because of the low gas prices paid by Carnation. A five stage, continuous fermentation process has been selected because of its lower costs and operating needs. A plant incorporating batch fermentation will require a price increase of $0.04/L to achieve comparable after-tax IRR.

A 25 percent increase in capital and operating costs will necessitate a $0.10/L price increase for both products and each scenario to maintain the after-tax IRR in the range of 25-35 percent.

Several companies wish to hold discussion with Carnation regarding potential investment and/or purchase of the product. Carnation does not wish to own and operate the ethanol plant, as it is not interested in entering the chemical/fuel business. However, Carnation should actively promote this project. Immediate actions should include support of the EM&R sponsored project into on-site, pilot-scale membrane work and laboratory fermentation tests. This will help resolve most of the technical uncertainties identified during this study. Carnation should solicit proposals from interested companies regarding construction, ownership and operation of the facility. The project viability can then be revised in October, 1988 and commitment to proceed made.

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INTRODUCTION

Ethanol has been described as one of the most exotic organic chemicals because of its unique combination of properties as a solvent, a germicide, a beverage, an antifreeze, a fuel, and as a feedstock for the production of other chemicals. A use of current interest is for octane enhancement and gasoline substitution. Ethanol is produced conventionally from molasses, grains and ethylene. Cost of the raw material is a significant component of the total ethanol production cost, and this is one reason why production of fuel ethanol from these conventional sources has not become widespread in Canada.

An opportunity exists to produce ethanol economically using waste materials at Carnation Foods' Carberry Plant. This major potato processing facility produces french fries and specialty potato products. The plant presently operates 5 - 7 days per week, 280 days per year. Solid wastes are hauled to a cattle feed facility and liquid wastes are treated in a lagoon. The objective of this study is to evaluate the use of solid materials after some preprocessing, and clean liquid streams following preconcentration using membrane technology, for the production of ethanol. Wardrop Engineering Inc. was commissioned by Carnation Foods; Energy, Mines and Resources Canada; and Manitoba Energy and Mines to explore this opportunity.

This report presents the results of our investigations into fuel and food/industrial grade ethanol production at the Carberry Plant. Energy efficient technologies such as mechanical vapour recompression and high temperature heat pumps have been considered for distillation. Use of waste starch from Old Dutch Foods' Winnipeg plant has also been evaluated, as it can increase plant capacity and sugar concentration without adding to the capital and operating costs proportionally.

Experimental and waste analysis data used for this study were generated through a joint effort between Carnation, Canadian Food Products Development Centre (CFPDC) and Wardrop. CFPDC, under a provincially funded project, collected blancher and cutter overflow samples, analysed these, and conducted bench-scale membrane separation tests. Wardrop, under the same project, developed a membrane system optimization model which was used for this study. Carnation's personnel collected samples for this study.

EXISTING OPERATIONS

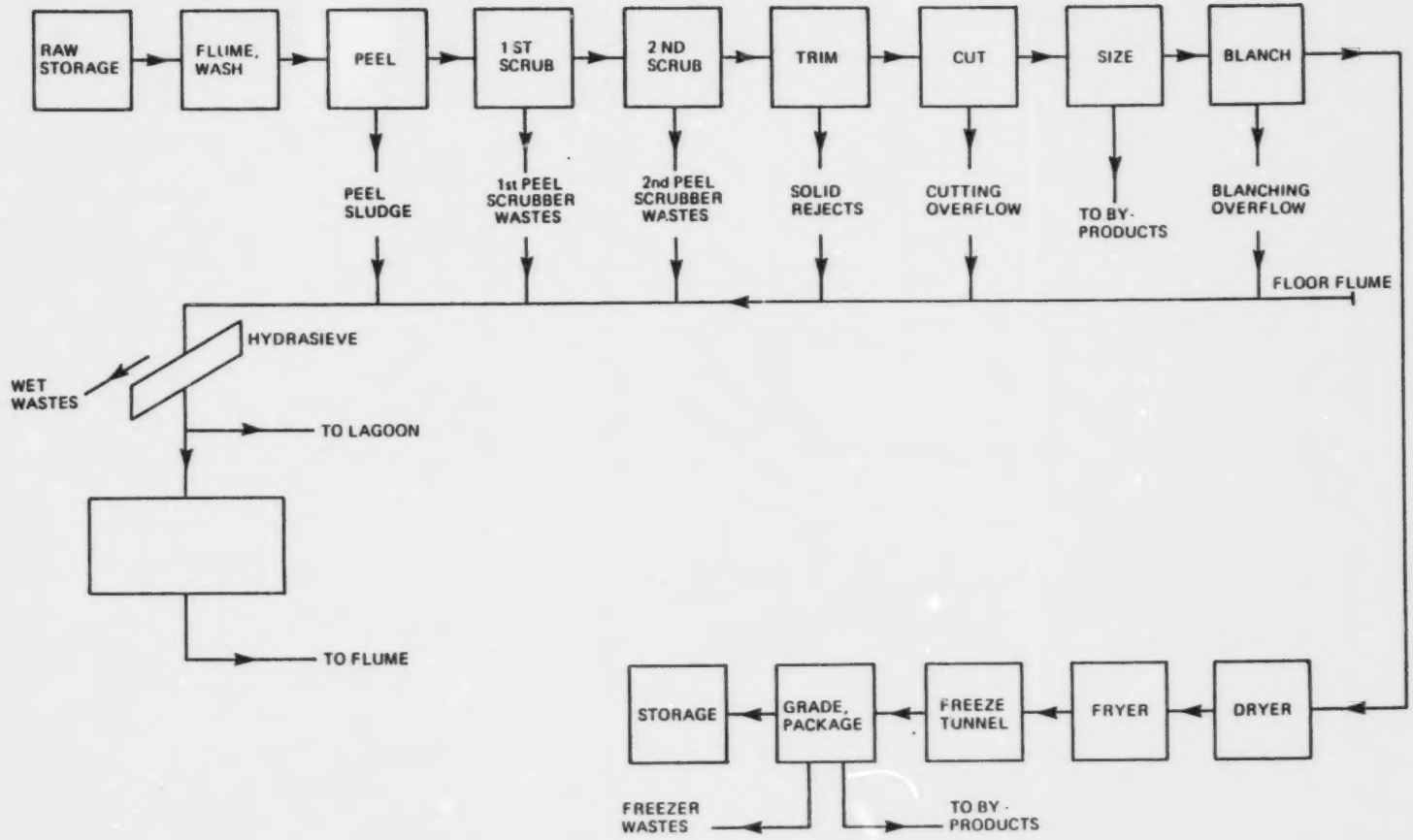
Carnation's Carberry plant produces frozen french fries and other potato products. Small potato pieces are processed into gems or MacDonald hash brown patties by the preform line. Even smaller pieces are processed by the granule line into mashed potato powder. However, a significant fraction of the raw material is wasted from the french fry process, and is available for ethanol production.

French Fry Process

Figure 1 is a block diagram of the french fry process. Raw potatoes are flumed from storage to the peeler room where these are washed and steam peeled. After peeling, two scrubbers remove loose peel and starch. In the trim room, workers remove eyes and blemishes. Then the potatoes are pumped to one of two cutting decks where these are cut into fries. Slivers are rejected in a vibrating screen and the rest is fed to the blanchers. In the blanchers, potatoes are heat treated to reduce enzyme activity, and leached with water to reduce sugar content.
After blanching, the fries are dried, par-fried, frozen in a freeze tunnel, graded and packaged.

As shown in Figure 1, all processing wastes are dumped into the floor flume, except for freezer wastes which are collected in bins and dumped outside. A hydrasieve removes all large solid pieces from the flume discharge and liquid that passes through the hydrasieve is either reused for raw potato fluming or wasted to the lagoon. Streams of interest for ethanol production include:

- Peel Sludge
- Peel Wastes
- Solid Rejects
- Cutting Wastes
- Blanching Overflow
- Freezer Wastes



FRENCH FRY PROCESS

FIGURE 1

Fresh process water is obtained from wells approximately 300 m north of the production plant. The boiler room has two boilers to provide high pressure steam for processing and for space heating. One has a capacity of 27,200 kg/hr and is operated continuously. The second is smaller, 9,000 kg/hr, and is used only when needed. It is expected that when running at full capacity in the winter, there will be little excess boiler capacity available for an ethanol plant.

Experience with Production of Ethanol from Potato Wastes

Simplot produces 34,000 L/d of ethanol from solid wastes at its Caldwell, Idaho, potato processing complex. This highly efficient plant has been operating successfully for many years. However, no dilute sugar bearing streams are used. Thus, the ethanol plant adds to the total organic load to the lagoons. It also uses small reject potatoes for ethanol production, which further adds to the organic loading. Carnation plant uses these small potatoes for specialty products or granule products.

The key difference between the Simplot project and the proposed Carnation project is that dilute sugar bearing streams would be used in Carnation's case after membrane concentration. This would result in an increased ethanol capacity and lower organic loads to the lagoons.

WASTE CHARACTERIZATION

The following streams are available for ethanol production:

- Peel Wastes
  - Peel Sludge
  - First Peel Scrubber
  - Second Peel Scrubber
- Cutting Water Overflow
- Blancher Overflow
- Wet Solid Wastes
- Freezer Wastes

A preliminary sampling campaign was carried out to characterize the above stream except the peel sludge which could not be sampled because of the location and the difficulty in getting representative samples. Results of this sampling campaign are presented in Table 1.

Grab samples were taken in all cases by Carnation and Wardrop personnel, except in one case where a 24 hour proportional blancher overflow sample was collected by the CFPDC personnel. All analysis work was performed at CFPDC. The total solids content was determined by weighing before and after drying, and glucose content measured by the glucose oxidase method.

This technique consists of the following steps:

- hydrolyzing starch in the sample into glucose with amyloglucosidase.
- measuring glucose concentration with a glucose oxidase titration.

Results are therefore reported as glucose equivalent carbohydrates. This glucose oxidase titration is highly accurate as error is estimated at $\pm$100 ppm (or 0.01 percent) for higher concentrations and 200 ppm (0.02 percent) for the lower concentrations. One possible concern is that this technique would not measure sucrose or fructose, but it is likely that these sugars are present at low concentrations in raw potatoes.

TABLE 1
WASTE CHARACTERISTICS

	Oct. 1986	Apr. 24, 1987	May 7, 1987	June 5, 1987	July 17, 1987	Flow (L/min)	Quantity (kg/d)
First Scrubber Wastes							
Carbohydrates, %[a]	-	2.3	1.2	1.4	0.8	45	
Solids, %	-	3.9	4.7	3.6	3.1		
Second Scrubber Wastes						280	
Carbohydrates, %	-	2.0	1.0	1.6	0.8		
Solids, %	-	3.8	1.8	2.8	1.2		
Cutting Wastes						460	
Carbohydrates, %	-	0.8	0.3	0.2	0.5		
Solids, %	-	0.6	0.6	0.3	0.6		
Blancher Wastes							
Carbohydrates, %	0.5[b]	0.2	0.0	0.2	0.1	600	
Solids, %	0.7[b]	0.4	0.6	0.6	0.65		
Wet Solid Wastes							
Carbohydrates, %	-	-	-	9.4	18.4		58,000
Solids, %	-	-	-	12.9	22.2		
Freezer Wastes							
Carbohydrates, %	-	-	-	34.5	30.4		48,000
Solids, %	-	-	-	44.2	38.4		

Notes: a - Carbohydrates are reported as glucose equivalent.
b - Data from Food Centre, based on a 24 hour cumulative sampling.

The flowrates for the liquid streams are normally constant and well known to the plant personnel. However, these were confirmed through bucket and stopwatch method, and found to be consistent with plant personnel data. Solid waste quantities were measured by weighing the haul trucks before and after loading the wastes.

Peel Wastes

Steam peeling results typically in losses in excess of 10 percent of the feed. This is conceivable as removal of 3-4 mm of the outer tissue along with the skin proper could represent a loss of over 30 percent of the total solids of a 225 kg tuber. For Carnation, peel losses are estimated at to 87,000 kg/day. Of this loss, the actual peel quantity is only 4,000-5,000 kg/day and the balance is potato pulp. For a starch content of 18 percent in raw potato, this represents a carbohydrate loss in the range of 14,000 to 15,000 kg per day.

The peel wastes are discharged from the following three process steps:

- Flash Steam Cyclone (peel sludge);
- First Peel Scrubber;
- Second Peel Scrubber.

The peel sludge stream could not be sampled as it is discharged at the bottom of the flume. It was also not possible to measure its flowrate. Modifications to the system are required to permit representative sampling and flow measurement.

The carbohydrate content of the first scrubber waste stream ranged from 0.87 to 2.3 percent. Variation likely occurred due to the variable peeler efficiency, and changes in the quality of the potatoes. Based on an average concentration of 1.4 percent and a flow of 45 L/min, approximately 1,350 kg of carbohydrates are discarded in this stream daily.

The second peel scrubber wastes contain partially hydrolyzed starch. This stream is very clean as nearly all the peel and dirt is removed by the first scrubber. Carbohydrate content was measured to be in the range of 0.5 to 2.0 percent. On a specific day, when the flow was measured to be 140 L/min for each peeling line, the concentration was 0.8 percent. Based on a

140 L/min flow for each line, at an average carbohydrate concentration of 0.8 percent, 3,200 kg of carbohydrates are available in this stream daily.

Cutting Water Overflow

Cutting water overflow carries away the starch granules released by cutting the raw potatoes. Average flow was measured at 230 L/min per line, but according to the plant personnel, is quite variable. Carbohydrate content ranged from .2 to .8 percent, with an average of .45 percent. It was also noted that most of this was in the form of insoluble starch granules, so the suspended solids content was also measured, as well as its starch content. It was determined that cutting wastes contain 0.4 percent filterable solids, which in turn contain 91.7 percent starch on a dry basis.

Thus we estimate the starch quantity that would be recovered by a hydrocyclone and sedimentation to be .4 x .917 = .37 percent solid starch. Therefore, the starch available for alcohol production from this stream is estimated at 2,400 kg/day.

Blancher Overflow

Blancher overflow contains soluble sugars removed from the raw potatoes during blanching. This is quite variable, as the free sugar content varies through the season. Also, the blancher is drained each day around noon, which reduces overflow sugar content for a while. Carbohydrate content was measured at 0 to 0.2 percent during this sampling program. However, the Food Centre had previously measured the carbohydrate content at 0.5 percent using a more accurate proportional sampling technique over a 24 hour period. Using an average concentration of 0.3 percent, and a flow rate of 600 L/min, the carbohydrate content is estimated at 4,150 kg/day.

Wet Solid Wastes

Wet solid wastes are solids removed from the flume discharge by a hydrasieve. These consist of mostly potato trimming wastes and some peel pieces, and have the consistency of raw potato. Wet wastes are presently hauled to a feed lot, and we requested the feed lot to monitor daily shipments. During a one month period, shipments averaged 58,000 kg/day. Carbohydrate content was measured to be 18 percent. During this sampling campaign, the plant was operating at 78 percent of its normal capacity because of start-up considerations. At the design capacity, carbohydrate rejected through this waste is estimated at 13,400 kg/day and is assumed to be available for ethanol production.

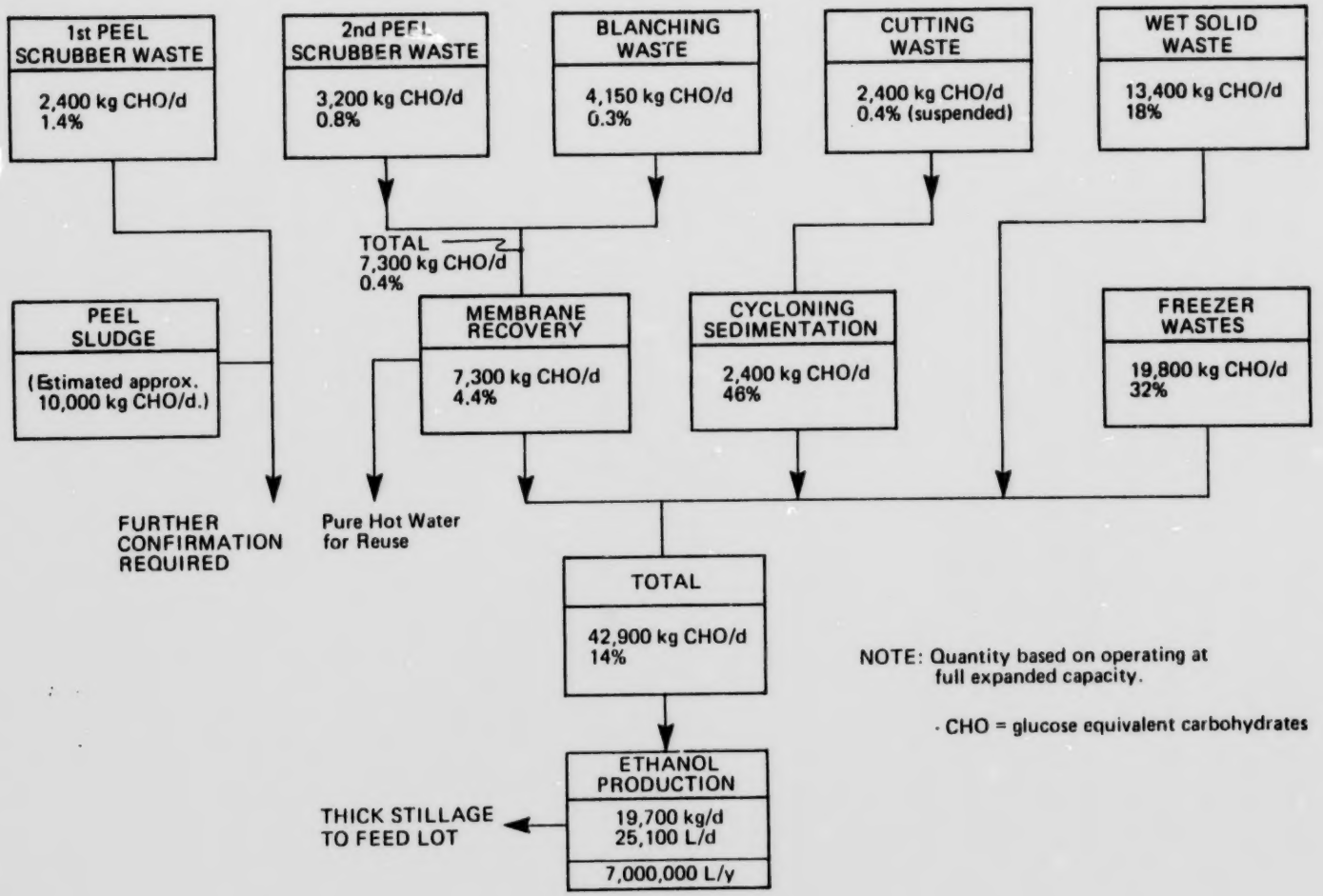
Freezer Wastes

Freezer wastes consist mostly of frozen fries and potato crystal rejects. These wastes are also shipped to the feed lot and were monitored there. Daily shipments averaged 48,000 kg/day and carbohydrate content averaged 32 percent. Assuming that the quantity of this waste will also increase by 28 percent next year under normal operating conditions, the quantity of carbohydrates available is approximately 19,800 kg/day.

Freezer wastes contain oil from frying. Its effects on milling and fermentation, are unknown. Concerns exist regarding the deposition of animal fats on pipe and vessel surfaces. Tests are required to determine the extent of these problems and solutions.

Total Wastes Available

Figure 2 summarizes the quantities of carbohydrates available from each waste stream. The total amount of carbohydrates available for alcohol production is estimated to be 42,900 kg/day. Based on 90 percent plant efficiency, the plant is expected to produce 25,100 litres of alcohol per day, or 7 million litres per year.



WASTE CHARACTERISTICS AND ETHANOL PRODUCTION

FIGURE 2

Although seven million litres/year has been assumed as the reference capacity, a number of factors will affect this:

1. Quantity and composition of peel waste: The peel sludge and first scrubber could increase plant capacity by as much as 15 percent.

2. Potential changes in capacity and operating schedule: Because of the recent increase in the plant capacity, it is currently planned to change operations to five days per week. This could reduce capacity by up to 15 percent. Also, some fluctuations in plant capacity occur depending on the availability of the raw material. For example, summer shutdowns may extent up to two months, but spanned only two weeks during the 1987 processing period.

It is thus desirable to test the impact of capacity fluctuation on project viability.

Another concern is the availability of wet wastes and freezer wastes to the ethanol plant, and its impact on the feedlot operation. This aspect is discussed in more detail in the Conceptual Design section.

MARKET ASSESSMENT

Two grades of product can be produced at Carnation's Carberry Facility:

- Fuel
- Food/Industrial

The fuel grade product is anhydrous (99.5 percent) ethanol suitable for blending with gasoline in 10 percent:90 percent ratio to produce gasohol. It can also be used as a co-solvent for methanol gasoline blends in 5 percent methanol, 3 percent ethanol and 92 percent gasoline ratios. The food/industrial grade product can be azeotropic or anhydrous, but meets the high purity pharmaceutical grade specification. It is suitable for vinegar production, for cosmetics and pharmaceuticals and as a solvent in a variety of uses.

A preliminary market survey has been carried out in both bases to determine potential markets and prices. Focus has been on determining the names of the companies which are interested in investing in the facility in a joint venture with Carnation while assuming the sole responsibility for using and/or marketing the product.

Fuel Ethanol

Current Activities

Mohawk Oil is involved in blending gasoline with ethanol and methanol/ethanol mixture. E-10 gasohol products, which contain 10 percent ethanol and 90 percent of various grades of gasolines, are sold in Manitoba. Experience with the use of this these products has been very positive and minimum problems have surfaced. As a result, the number of Mohawk gas stations in the Province has expanded, and sales have increased.

Mohawk currently produces approximately seven million litres per year of ethanol at its Minnedosa plant from grain. Because of the high cost of production from a small facility, a provincial road tax relief of $0.025/L of

gasohol provides a subsidy of $0.25/L of ethanol. Mohawk is currently expanding production to approximately 8.5 million litres per year by adding two fermenters.

In addition to the marketing of E-10 gasohol in Manitoba, Mohawk has also pioneered a methanol-ethanol-gasoline (EM) blend. This is being marketed in Saskatchewan, Alberta and British Columbia. The ratio is five percent methanol, three percent ethanol and 92 percent gasoline. The low cost of methanol is the key economic driving force in this case which permits the use of more expensive ethanol. The September, 1987 list price of methanol in the U.S. was $0.36-0.38 U.S./U.S. gallon. This translates into a Canadian price of approximately $0.13-0.14/L. Normally, prices in Canada are $0.01-0.02/L higher than the equivalent U.S. price. It is estimated that this would result in a value of $0.40/L for ethanol based on $0.215/L refinery gate gasoline price.

In Alberta and Ontario, provincial road tax is forgiven on the alcohol portion of the blends, resulting in a further cost advantage of $0.08/L of ethanol in Alberta and up to $0.14/L in Ontario. However, methanol-ethanol blend is not being marketed by Mohawk in Ontario because of the volatility restrictions defined by the Canadian General Standards Board. This issue is currently under CGSB review, and the potential for change in these regulations looks bright. If a change favourable to ethanol-gasoline and ethanol-methanol-gasoline blend occurs, a large market will open up in Ontario for ethanol.

Currently, two companies, V-Plus and Pioneer, are marketing methanol-isopropanol blends in Ontario. These firms could use ethanol as a co-solvent as in this case, gasoline is being modified in the refinery to permit compliance with volatility regulations. We also understand that Suncor is planning a two percent methanol - two percent ethanol blend in a limited number of stations in Ontario.

Recently, Saskatchewan has announced a generous ($0.045/L) tax relief on E-10 gasohol. This translates into a $0.45/L advantage for ethanol. However, this incentive applies to Saskatchewan produced ethanol only. This could lead to production activities, as well as marketing of E-10 blends in that province.

Value of Fuel Ethanol

Based on the current prices of methanol and gasoline, it would appear that an ethanol price in the range of $0.35-0.40/L will be attractive for E-M blending. A base price of $0.30/L should be used for an economic analysis.

The value of ethanol will be substantially higher for E-10 blending in Manitoba ($0.50-$0.60/L). This subsidy was provided to help production of ethanol from biomass, and remains available to potential producers. As reported in Biofax (No. 17, 1987), a Finance Department spokesman has stated that this subsidy will likely be reviewed if oil prices reach $30 US/barrel.

Discussions with Potential Users

A number of potential fuel ethanol users were contacted to determine their interest in participating in the proposed plant or purchasing ethanol from it. These include:

- Mohawk Oil: Mr. Donald O'Connor, Manager - Gasohol Division. Mr. O'Connor is very interested in holding further discussions with Carnation regarding joint ownership/ethanol purchase, if the production prices are in the range of $0.25-0.30/L.

- Alberta Gas Chemical: This company markets methanol/isopropanol blend in Ontario through the V-Plus outlets. They can blend ethanol and still meet volatility restrictions as the gasoline is adjusted in the refinery to reduce volatility. AGC appears to be interested in holding further discussions with Carnation if the price is $0.30/L, but does not know if interest in participating in the project would exist.

- Petro Canada: Petro Canada would like to talk with Carnation further but our contact was not very optimistic. The size of the ethanol production capacity at Carberry is too small for Petro Canada, if the company were to launch into ethanol. The best place to blend is at the refinery, and Ontario will be the logical place because of the road tax relief on alcohols. This will result in an ethanol value of $0.28-0.30/L.

- Suncor: Suncor is now participating in the blending program with AGC. However, they still have not committed themselves to an octane enhancer. Suncor would like to explore the matter further if the selling price is in the range of $0.25-0.30/L.

Food/Industrial Grade Ethanol

Ethanol has numerous industrial uses. These range from solvent for printing inks to cleaning, antifreeze and toilet preparations. However, these markets are relatively small, require expensive packaging and have well established suppliers in place.

Of a greater interest is the vinegar market. Azeotropic ethanol of high purity is used as a raw material for vinegar production. In 1984, 50,443 cu.m. of spirit vinegar was produced in Canada. This is equivalent to 6,000 cu.m. of pure acetic acid, and approximately 7,000 cu.m. of pure ethanol, equal to the potential capacity of the proposed ethanol plant. This market is currently being supplied by Commercial Alcohols Ltd. of Montreal, which uses ethylene as the starting material. Major vinegar producers include Heinz, Rheinhart and Fleischman (who recently purchased the Canada Vinegar label from Campbell Soup). The bulk of Heinz production is for in-house use, and Rheinhart in the largest producer of industrial vinegar. Only Canada Vinegar has a plant in the West, in Saskatoon, which uses 20,000 L of ethanol/month.

Fleischman is a major producer in the U.S. It uses 34 million litres of ethanol annually for vinegar production.

Although none of the vinegar producers released the price they are paying for ethanol, it fluctuates with the price of oil, and is estimated to be in the range of $0.45-0.50/L of pure ethanol. This is substantially lower than the list prices of $0.70/L. In the U.S., the list price of ethanol is $1.10/U.S. gallon or approximately $0.39 Canadian per litre. It appears that actual selling price is slightly above $1.0/U.S. gallon or $0.36/L Canadian. Subtracting a duty of three percent, this results in a value of approximately $0.35/L for vinegar grade ethanol for export to the U.S.

We discussed the possibility of purchasing ethanol or participating in the project on a joint venture basis with the major vinegar producers. In each case, interest existed in purchasing ethanol if the price was acceptable.

Conclusion

Markets exist for both fuel and food grade ethanol produced at the Carberry site if FOB plant price is in the $0.30-0.35/L range. Fuel ethanol market could expand rapidly because of the road tax relief if volatility regulations are revised in Ontario.

Two industrial ethanol consumers and two fuel ethanol users are interested in holding discussions with Carnation for the proposed facility and purchasing ethanol. Three of these four companies indicated an openness to discuss joint ownership of the project with Carnation. Additional companies may show interest if CGSB revises the volatility regulations to open up the Ontario market.

TECHNOLOGY ASSESSMENT

Since the oil crisis of the early seventies, interest in ethanol as an octane booster and a gasoline substitute has intensified all over the world. New technologies have been investigated and developed including acid hydrolysis of starch, continuous fermentation of grain-based substrates and more energy efficient distillation. For this project, an area of specific interest is the pretreatment of liquid waste streams to either recover starch or to concentrate sugars. This is aimed at increasing the plant capacity and reducing organic loading to the lagoon so that additional loads from the ethanol plant may be accommodated.

Preconcentration of Dilute Wastes using Reverse Osmosis

For economic alcohol production, the carbohydrate source concentration should ideally be at or above 14 percent. Fermenting dilute substrates results in cost penalties because as ethanol concentrations decrease:

- Distillation energy requirements increase.
- Fermentation and distillation equipment sizes increase.

Using this criteria, the blancher and peel scrubber wastes would normally be rejected. However, these streams can be concentrated at an acceptable cost using the reverse osmosis technology to make these suitable for cost-effective ethanol production.

The CFPDC had a project entitled "Evaluation of Commercial Applications of Separating Techniques to the Food and Biotechnology Industries". As part of this study, the concentration of some of Carnation's wastes was tested using reverse osmosis (R.O.). It was found that R.O. could concentrate the waste streams from 0.8 percent up to four or five percent solids at high flux rates.

These concentrations are high enough so that when mixed with the wet wastes, freezer wastes and cutting waste starch, a mash containing 14 percent carbohydrates can be obtained.

As part of the Food Centre's project, Wardrop wrote a computer model to simulate and optimize membrane separation systems. This model requires the following inputs:

- feed flow and concentration;
- desired product concentration;
- flux rates as a function of concentration;
- membrane cost;
- constraints on pressure, flow, concentration imposed by the membrane;
- membrane life;

Based on the above data, the model designs an optimized membrane system and estimates its cost.

Wardrop's R.O. system optimization model was used to design an R.O. system to concentrate the blanching and peel scrubber wastes. Results of the model are discussed in the Conceptual Design chapter.

Work at CFPDC and our analysis has provided the confidence that sugars can be recovered from the two dilute waste streams of interest and permeate of a quality acceptable for recycle can be obtained. However, on-site pilot tests are required to confirm design parameters, performance, and integration of the membrane system into the french fry operation. These tests will also confirm composition of the various waste streams.

Cutting Wastes - Hydrocyclone and Sedimentation

Most of the carbohydrate in the cutting wastes are in the form of insoluble starch granules. These starch granules would quickly foul an R.O. system. Therefore, hydrocycloning followed by sedimentation has been selected. This is an established method of recovering starch granules at up to 50 percent concentration in a cost-effective manner.

In this method, the starch stream is pre-concentrated to five percent solids in hydrocyclones. The potato starch granules have a specific gravity of approximately 1.5, and are in the 30-60-100 micron size range and settle rapidly. A 50 percent solids concentration is achievable in a clarifier with a 30 minute residence time.

Starch Hydrolysis

The conventional process for converting starch to fermentable sugars involves cooking followed by a two stage enzymatic reaction. In the first stage, a temperature resistant variety of alpha amylase is used which randomly disrupts the linear glucose molecule chain. A second enzyme, amyloglucosidase, converts these large oligomers into fermentable maltose and glucose. The bulk of this reaction is carried out at 60°C, although it continues during fermentation. This well established process has been assumed for this study.

St. Lawrence Reactors of Mississauga markets an acid hydrolysis process that replaces cooking and alpha-amylase reaction with a single step. However, amyloglucosidase must still be added to complete the hydrolysis to fermentable sugars. This technology might be competitive with the established technology and should be considered during the process design stage of this project.

Fermentation

The conventional method of converting sugar to ethanol is batch fermentation using a yeast strain such as Saccharomyees Cerevisiae as the fermenting microorganism. Although S. Cerevisiae remains the most commonly used microorganism, others such as strains of Zymomonas Mobilis have been tried. However, the most significant single innovation in fermentation has been the introduction of continuous fermentation with or without yeast recycle.

Continuous fermentation has been used in the production of ethanol from sulphite liquor for many years. The key reason is that the presence of sulphur dioxide helped control contamination. As the sulphite liquor is clear, yeast can be removed through centrifugation and recycled. This process was used at the CIP's Gatineau plant which stopped ethanol production when the pulping process change reduced the amounts of fermentable sugars present in the liquid.

A more recent innovation in continuous fermentation is the Biostill process developed by Alfa Laval for molasses fermentation. This process is designed for concentrated feedstocks, and involves a single stage fermentation at low sugar and high ethanol concentrations. Yeast is recovered through centrifugation and recycled. Although the process has been extended to grain fermentation, the objective remains to minimize the use of water. Biostill is not suitable for the proposed application because of the dilute nature of Carnation's wastes.

Vogelbusch, an Austrian company, has developed a continuous, multi-stage fermentation process suitable for the proposed Carnation plant. For grain fermentation, yeast recycle is not considered possible because of the presence of insoluble non-fermentables. Partial yeast recycle may be possible at Carnation depending on the level of such insolubles. The process involves five fermenters connected in series. Prefermenters continuously generate yeast, which is used to inoculate the first fermenter. Advantages of the process include ease of operation and reduced tankage requirements. A potential disadvantage is the risk of contamination although the process has been successfully used at a 200 million litres per year fuel ethanol plant at Pekin, Illinois for over three years with minimal contamination problems. Vogelbusch also claims similar successes with the process in Europe and Africa.

For this project, we have used Vogelbusch's continuous fermentation process for the base cases. However, conventional batch fermentation has also been evaluated for comparison.

Distillation

In the conventional distillation process, energy is supplied in the form of steam. This technology will be used for formulating the base cases. The following energy efficient ethanol distillation technologies are discussed below:

- Heat Pump Powered Distillation
- Mechanical Vapour Recompression Distillation

Heat Pump Powered Distillation

Conventional alcohol distillation requires both heat supply and removal. As shown in Figure 3, a reboiler supplies heat to the bottom of the column to drive the more volatile material, alcohol, up the column. The reboiler is usually heated with steam. (Alternatively, live steam can be injected into the column base.) The alcohol concentrates at the top of the column where it is condensed in a water cooled condenser. Some of the condensate is returned to the column as reflux while the balance is removed as product.

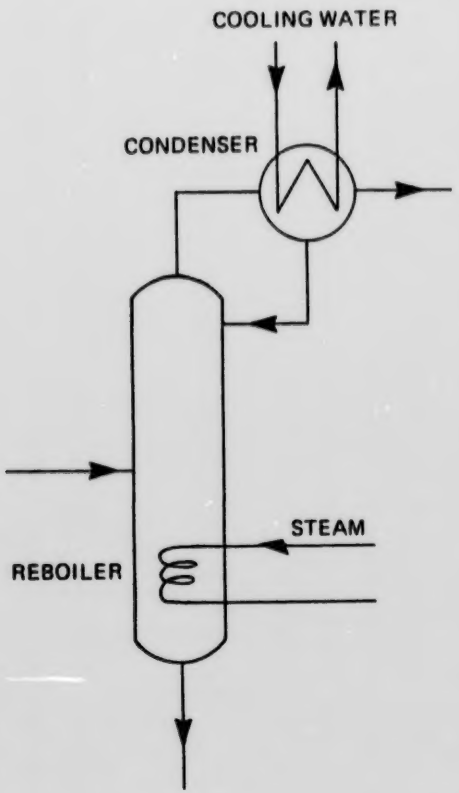
Because both heat supply and removal are required, distillation is an ideal candidate for using a heat pump. As shown in Figure 4, a refrigerant loop is used to return heat from the top of the column to the bottom. Because the bottom is always hotter than the top, heat pumping is required. A compressor causes the refrigerant to condense in the reboiler at a temperature higher than the boiling point of water. COP's are usually in the range of 3 to 5, i.e. for every kilowatt of electrical power used by the compressor drive, 3 to 5 kilowatts of heat are supplied to the reboiler.

For ethanol distillation, heat is required at 110°C or more to boil water. Standard heat pumps using the common refrigerants such as R12 or R22 cannot provide heat at these temperatures. Heat pumps using R114 that can provide heat up to 110°C are commercially available, although only a few are in operation. Other refrigerants, such as R113 and perfluoro-n-hexane, are also being considered, but are behind R114 in development.
Several phone calls to high temperature heat pump suppliers were made to determine the status of this technology. Apparently there is one steam generating HTHP's operating in the U.S. and two in Canada. These are:

- A plant in the U.S. Rocky Mountain states
- NL Chemical Plant, Varrenes, Quebec
- CIP Newsprint Mill, Gatineau, Quebec

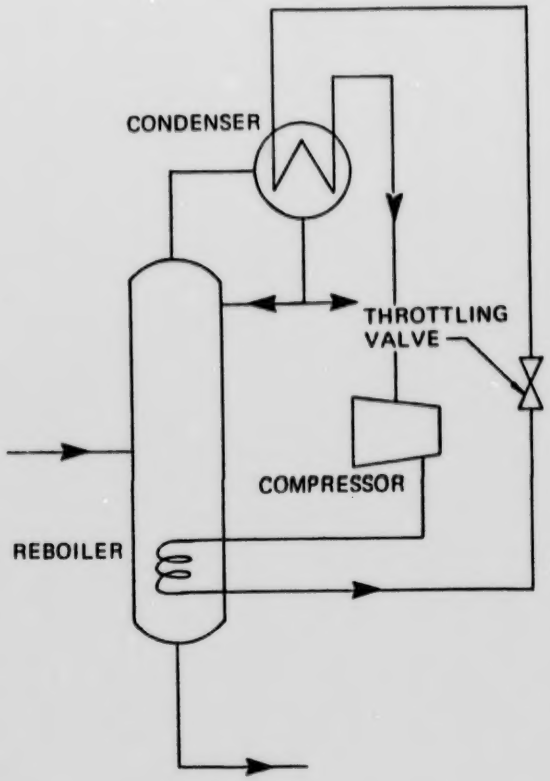
However, these are still relatively new and the technology is not risk free. None has been applied to ethanol distillation, although York claims to have used HTHP's for hydrocarbon distillation.

FIG. 3



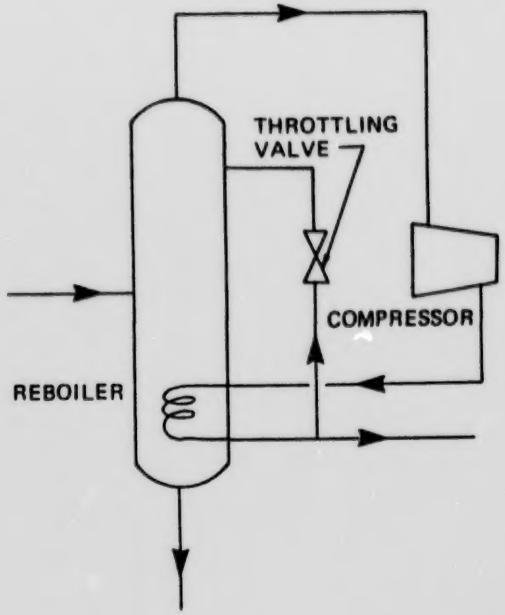
CONVENTIONAL DISTILLATION,

FIG. 4



HEAT PUMP
POWERED DISTILLATION

FIG. 5



MECHANICAL VAPOUR
RECOMPRESSION DISTILLATION

A high temperature heat pump has been evaluated for the proposed ethanol plant and compared with conventional steam distillation.

Mechanical Vapour Recompression Powered Distillation

An option similar to a high temperature heat pump is to use mechanical vapour recompression (MVR) to power distillation. A simplified diagram is shown in Figure 5. In this process, overhead alcohol vapours are compressed to increase their condensation temperature to 110°C or more. Alcohol condensing in the reboiler heats the distillation column. Condensed alcohol is throttled to atmospheric pressure before product removal or refluxing.

MVR has a few advantages over high temperature heat pumps:

- Only one large heat exchanger is required for MVR distillation. Two are needed for heat pump distillation.

- Significantly higher COP's are obtained for MVR, in the range of 8 to 10 as compared to 3 to 5 for heat pump.

- No refrigerant is required.

Sulzer Brothers Ltd. is the world's leader in MVR distillation. Their technology is available through Sulzer Canada Inc. or through Koch Engineering Company Inc., who are licensees of Sulzer Brothers. Koch have applied MVR to hydrocarbon distillation and claim that they have tested MVR for alcohol in a large pilot unit.

A significant part of the cost of either Sulzer or Koch's MVR distillation system appears to be special low pressure drop column packing. High column pressure drop increases the temperature difference between column top and bottom which decreases COP. Thus, to obtain a high COP, they specify using their low pressure drop packing.

This system will be evaluated further in the report.

Dehydration

The conventional technology for converting azeotropic ethanol-water mixture to anhydrous ethanol consists of distillation with an entrainer such as benzene or cyclohexane. A recent innovation is the use of zeolites for dehydration. A vapour phase molecular sieve zeolite separation process uses only 10 percent of the energy required for conventional dehydration, and is cost-effective. It is easier to operate than the more conventional processes which require careful control of entrainer concentration. Also, the entrainer must be constantly recovered, purified and recycled.

Based on these considerations, molecular sieve dehydration is used as the reference process for this report.

Another process of interest is the use of membrane technology in a configuration called pervaporation. This process is being offered commercially by Vogelbusch, but is not considered in this report because of its recent introduction.

CONCEPTUAL DESIGN

This section presents the description of the process and facilities, and discusses plant operating needs and practices.

Mass Balance

Appendix A presents a preliminary mass balance for the facility. Production rates for products and by-products are as follows:

Ethanol	= 7 million L/year
By-Product Feed	= 15,000 kg/d (10-15 percent solids)
Fusel Oil	= 50,000 L/y

Both fuel and food grade alcohol production produce stillage as the main waste product. Stillage contains unfermentable materials from the potato feed, such as cellulose, as well as dead yeast. A 100 mesh vibrating screen removes a portion of the insoluble material, and the balance, called thin stillage, is discharged to the sewer. Screen solids, called thick stillage are estimated to be in the range of 15,000 kg/day at a protein content of 10-15 percent. However, fermentation tests are required to confirm this.

Thin stillage will likely add to the total organic load on the lagoons because of the unfermentable organics present in the wet solid and frozen waste streams which are currently being diverted to the feed lot. However, this increase is expected to be small as soluble sugars in peel and blancher wastes will be diverted from the waste streams and fermented. Changes in organic loading to the lagoon can be best estimated through fermentation of various streams, removal of ethanol and measurement of organic loading in the stillage.

In fuel grade distillation, fusel oil removed by the fusel oil separator can be added back to the product upstream of the dehydration unit. Alternatively, fusel oils can be sold separately to organic chemicals producers, if a higher price is available.

In food grade distillation, a heads draw byproduct is produced along with the fusel oil. These cannot be mixed back with the product and should be sold separately. The heads draw can be reprocessed on a day when the distillation equipment is available to recover its ethanol content.

In the financial analysis, no value for any of these byproducts was assumed for all options. Any profit from byproducts would be small and variable. Also, no cost for disposing of stillage was included as the waste disposal costs at the Carberry plant are small.

Process Description

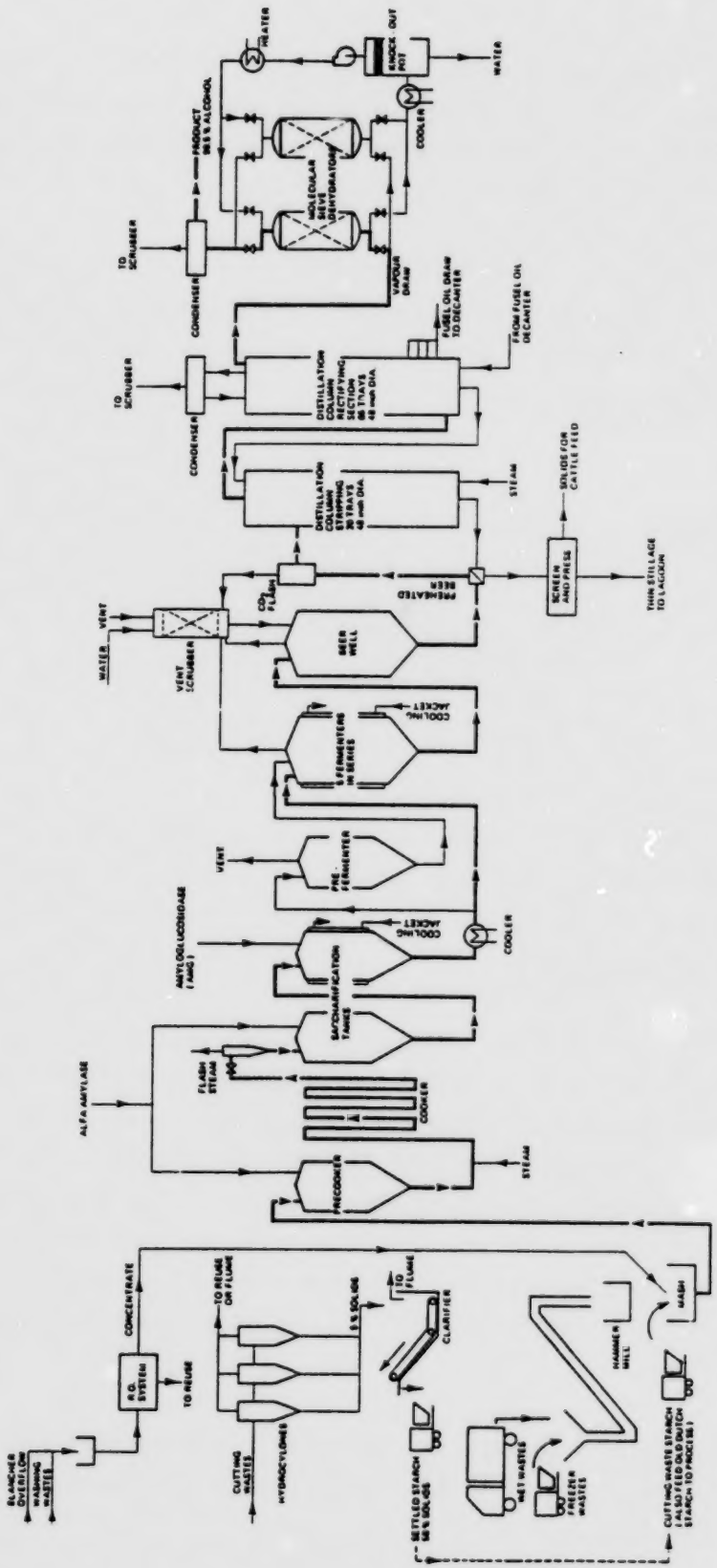
Figure 6 presents a process flow diagram for fuel grade ethanol production. Description of each process step follows:

R.O. System

Blanching and peel scrubber wastes are combined and concentrated by the reverse osmosis system. Mixed feed concentration of 0.5 percent and carbohydrate levels of 3.4 percent in the concentrate stream are used for designing the membrane system.

Wardrop's membrane optimization computer model was used to prepare a preliminary design of the R.O. system. Filmtec spiral wound membrane modules were used in this analysis as these are compact and relatively low in cost per unit membrane area. Based on the Food Centre's data on flux rates, and cost information collected by Wardrop and the Food Centre, the optimum design would require 420 m^2 of membrane in the form of 14 modules, each containing 30 m^2. The desired concentration factor of 7 is obtained in a single pass through the membrane at an inlet pressure of 4,100 kPa. Flow rate of the mixed feed stream is 550 L/min.

The R.O. system would be located in the french fry production area. R.O. permeate is expected to be of an acceptable quality for reuse in the blanching and scrubbing processes.



PROCESS FLOW DIAGRAM FOR BASE CASE FUEL GRADE ALCOHOL **FIGURE 6**

Hydrocyclone - Clarifier System

Starch granules in cutting wastes are collected and concentrated by a battery of hydrocyclones and a clarifier. The wastes are pumped from both cutting decks to three small cyclones, which concentrate the starch to 5 percent. This concentrated stream flows to a clarifier where the starch settles to a sludge with a solids content as high as 50 percent. This is then transferred to the ethanol plant in large plastic containers.

Hydrocyclone overflow stream is expected to be suitable for reuse in the cutting loop. Clarifier overflow would probably be discarded to the flume.

Solid Waste Handling

Hauling solids wastes from their sources to the ethanol plant is the most straightforward transfer method.

Freezer wastes are now collected in small bins in the packaging area and are dumped into a pile in the parking lot. These are hauled from there to the feed lot. The forklifts could dump the freezer wastes instead into a screw conveyor that feeds the hammer mill in the ethanol plant.

Wet wastes are now discharged from the hydrasieve into a truck to be taken to the feed lot. The same truck could be used to bring these to the ethanol plant to be dumped into the screw conveyor. Or, small bins and a forklift could be used as with freezer wastes.

Cutting waste starch, concentrated by the clarifier, would be difficult to pump. This starch should also be hauled to the ethanol plant using watertight bins and a forklift. As starch granules need not be milled, these can be fed directly to the mash tank downstream of the mill.

Pumping of mixed solid wastes is an alternate method of transfer. Piston ram pumps that can pump materials such as meat scraps are on the market and could be used to pump all of the wastes, mixed together, from the production building to the ethanol plant. Costs and ease of operation for these two

options should be compared during detailed design. This comparison cannot be finalized now as alcohol plant site and layout would influence it.

Once the wastes are hauled to the ethanol plant a screw conveyor feeds them to a hammer mill with 6 mm screen opening. Fresh raw potato is relatively easy to mill. If allowed to sit for several days, it can become rubbery and difficult to mill. Pilot tests should be done milling frozen french fries, but problems are not anticipated.

Concentrate from the R.O. system and the hydrocyclone - clarifier system are mixed with the solid wastes after milling. This mash should have a carbohydrate concentration of at least 14 percent.

Cooking, Liquefaction and Hydrolysis

Cooking and liquefaction are the first two steps in converting the starch to sugar. Mash is pumped to the precooker continuously, where it is heated to 60°C and a small dose of high temperature resistant alpha amylase is added. If necessary, the pH is adjusted to an optimum level for alpha amylase reaction.

Then the pre-cooked mash is pumped to the cooker. High pressure steam is added to the mash to heat it to 110°C. As the temperature increases, gelatinization occurs. Although bulk of the alpha amylase is destroyed as the temperature rises, liquefaction reaction occurs to a degree that acceptable mash viscosities are maintained, avoiding plugging of the cooker. Residence time in the cooker is approximately five minutes.

The mash is released from the cooker through a back pressure valve. This permits some water to flash, cooling the mash to 90°C. The main dosage of alpha amylase is then added to the first saccharification tank to continue starch hydrolysis.

The mash is pumped to the second saccharification tank, which is cooled with a cooling jacket to 60°C. Amyloglucosidase (AMG) is added, which is an enzyme that breaks the short starch molecules all the way down to maltose and glucose. However, AMG is inhibited by glucose, and only 70 percent of the starch is converted. The reaction is completed in the fermenter as glucose is fermented to ethanol.

Fermentation

In fermentation, Saccharomyees Cerevisiae converts the glucose produced during the hydrolysis step into ethanol and carbon dioxide. As the optimum temperature for fermentation is 30°C, the mash is cooled before fermentation, and also during fermentation to remove the heat of reaction.

Some of the cooled mash is pumped to a prefermenter, which is aerated to supply oxygen to achieve rapid yeast growth. (This consumes some raw materials but is a required cost.) Dry yeast is added to the prefermenter at start-up only. Nutrients such as ammonium sulphate and phosphate are also added to promote yeast growth.

The prefermented mash is added back to the rest of the mash which is pumped to five fermenters in series. Total residence time in the five fermenters is 48 hours. Any alcohol that evaporates and leaves the fermenter with the CO_2 vent is recovered by scrubbing with cold water in a vent scrubber. The first two fermenters have cooling jackets.

After fermentation, the fermented mash, called beer, contains 7 percent alcohol by weight, or 8 percent by volume. Beer is stored in a surge tank from where it is continuously fed to the distillation system.

Distillation

Figure 6 shows the distillation process required to produce an azeotropic ethanol water mixture. This system removes impurities from ethanol to an extent where the product is suitable for fuel use. Although one column could be used, two shorter ones are used to reduce headroom requirements. In the first called the stripper or beer still, ethanol is stripped off the beer in approximately 50 percent by volume concentration. Azeotropic ethanol is produced in the second column, called the rectifier.

The stripping section consists of approximately 20 sieve trays. Sieve trays are used because these tolerate solids well without plugging. The preheated beer enters the column at the top of the stripping section. Live steam is added at the column bottom to strip nearly all the alcohol from the beer and to

drive it up the column. The spent beer, called stillage, exits the bottom of the column, is used to preheat the beer feed and is then screened and sewered.

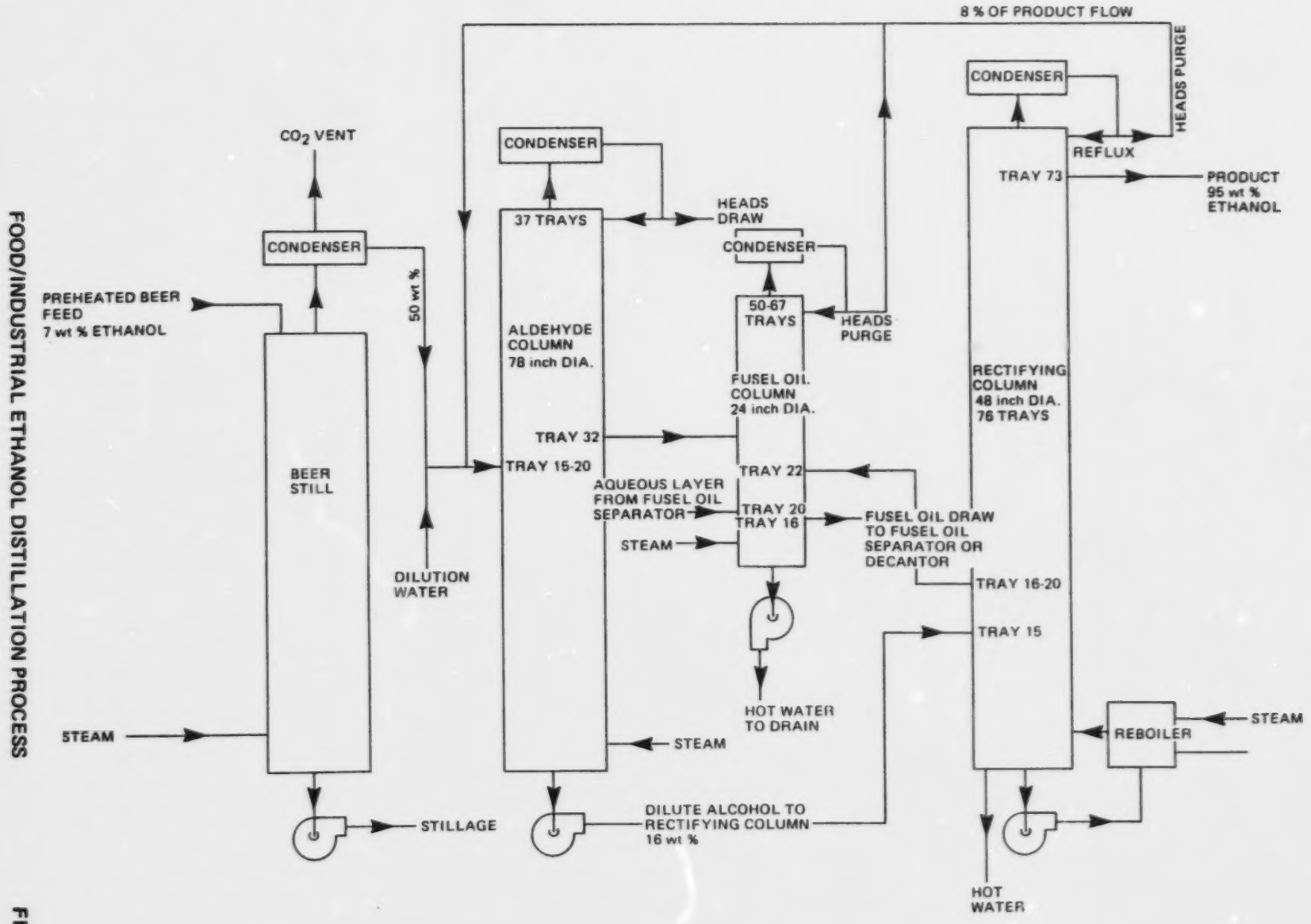
The rectifying section contains 65 bubble cap trays, although sieve or valve trays can be used. In this column, the ethanol-water vapour mixture enters the bottom at approximately 50 percent level. The concentration of ethanol in the vapour phase increases as it moves up the column through mass transfer with the down coming liquid on successive trays. A water cooled condenser condenses 80 percent of the azeotropic vapour at the top of the column which is returned to the column as reflux. The balance of the vapour is fed to the dehydration system for conversion to anhydrous product.

A byproduct of yeast fermentation is fusel oil, a mixture of higher alcohols and esters. The fusel oils are withdrawn from the rectifying column at a location where ethanol concentration is in the range of 60 to 70 percent by volume. Water is added to the fusel oil draw after it is cooled. This dilutes the alcohol concentration making the fusel oils insoluble. A decanter is used to separate the fusel oils from the aqueous stream. The fusel oil stream is approximately 0.8 percent of the product stream and contains higher alcohols, ethanol and water. The aqueous stream contains ethanol and is returned to the distillation column for recovery.

Dehydration

As discussed in the previous chapter, the fuel alcohol market requires alcohol containing less than 0.5 percent water. However, distillation can concentrate alcohol only up to the azeotropic 95 percent mixture. Dehydration is required to remove the remaining water and molecular sieve dehydration process has been selected for this application.

Union Carbide markets a molecular sieve dehydration system called the AHR Process, shown in Figure 6. Azeotropic vapours flow through one of the two adsorbent beds which removes practically all the water through adsorption. Most of the heat of adsorption goes into warming the molecular sieve bed. Dry alcohol vapour passes through the bed and is condensed by a product condenser. After a preset time, valves permit switching of the beds to connect the second bed to the distillation system and the first one to a regeneration system.



FOOD/INDUSTRIAL ETHANOL DISTILLATION PROCESS

FIGURE 7

Figure 6 shows the left bed on line with the process and the right one undergoing regeneration.

During regeneration, an inert gas, such as CO_2, is heated and passed through the bed to remove the moisture. The stored heat of adsorption helps in water desorption improving the overall energy efficiency. A cooler then cools the CO_2 to cause most of the moisture to condense and the drier CO_2 is recycled to the bed. The higher the dryness of carbon dioxide being returned to the bed, the lower is the bed's residual moisture content, and the higher is its ethanol dehydration efficiency. Because the Carberry plant has cold well water for cooling with temperatures of 5 - 7°C year round, alcohol with a moisture content below 0.4 percent can be produced.

Experience shows that the high temperatures encountered during desorption degrades the zeolite with time. A cost allowance to replace the zeolite regularly has been included in operating cost estimates.

Food/Industrial Grade Distillation

Food/industrial ethanol must meet more stringent quality requirements than the fuel alcohol. Thus the level of impurities, also called congeners, should be very low. Congeners include:

- fusel oil (mostly amyl, butyl and iso-propyl alcohols)
- aldehydes
- esters (light esters are mostly ethyl acetate, high boiling esters, iso-amyl acetate)

Figure 7 shows one distillation process that can produce alcohol with very low levels of congeners. This process uses three columns besides the beer still. The beer still concentrates the alcohol to a 50 percent ethanol solution called wine. The wine is fed to the middle of the aldehyde column along with some dilution water. Aldehydes and light esters (called heads) are more volatile than ethanol in this column because of their relatively low solubility in dilute alcohol solutions, and concentrate as they reach the top of the aldehyde column. A draw at the top of the column removes the heads continuously. Dilute alcohol leaves the column bottom and is pumped to the

rectifying column. A small stream also flows from the column middle to the fusel oil column to remove most of the fusel oils.

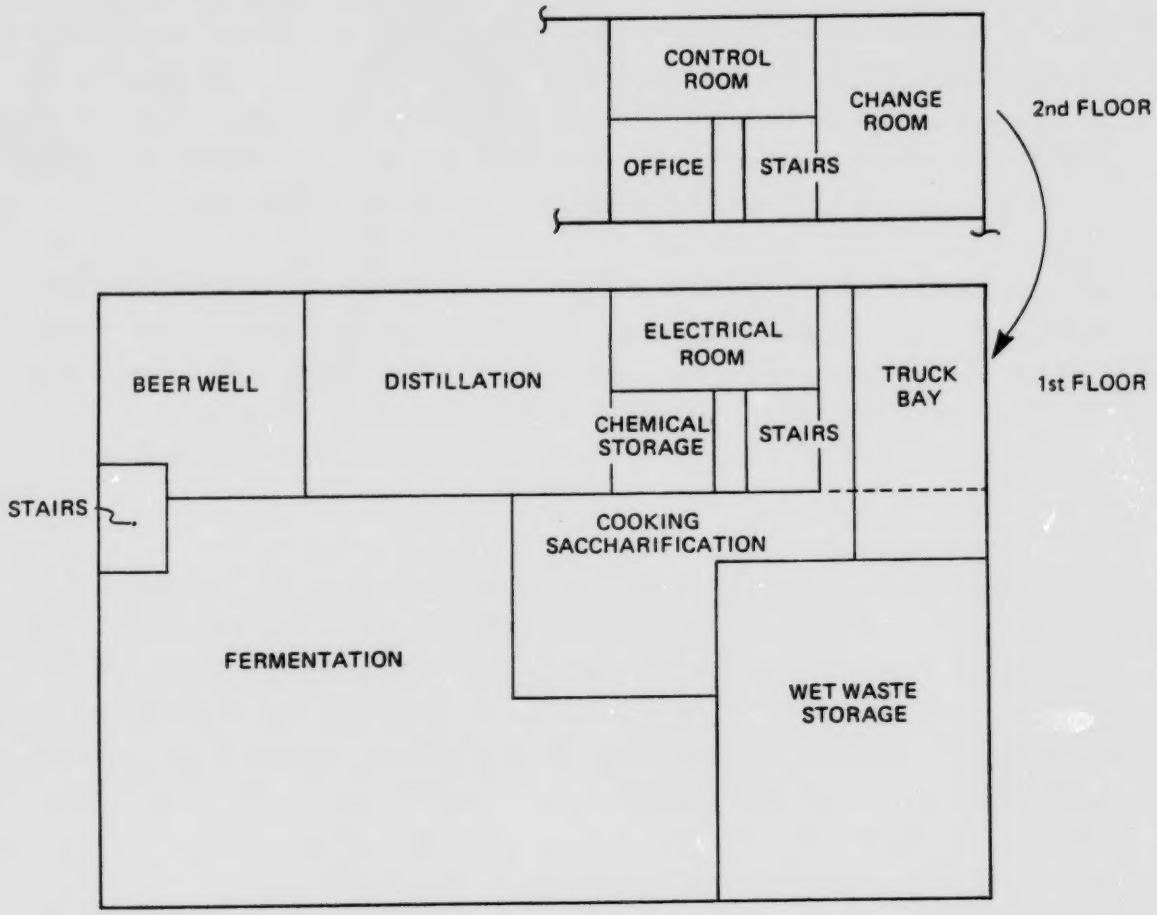
Fusel oil bearing streams are removed continuously from the aldehyde and rectifying columns. To meet stringent product specifications, flow rates for these sidestreams are much higher for the food/industrial product than those for the fuel product. As these streams contain significant quantities of ethanol, further processing takes place in the fusel oil column to recover and recycle ethanol to the rectifying and aldehyde columns. In the fusel oil column, a sidedraw is decanted to purge fusel oils and the ethanol stream recycled to the fusel oil column. Heads from the fusel oil column are pumped to the aldehyde column, and the aqueous bottom stream is sewered.

Dilute alcohol from the aldehyde column is concentrated in the rectifying column. A fusel oil draw removes any remaining fusel oil and a heads purge from the column top removes any remaining heads. If everything is operating properly, the heads concentration in this stream should be quite low, resulting in a very slight odour. Product is drawn three trays below the column top.

All four columns are constructed of 304 stainless steel. To reduce column heights, the rectifying and fusel oil columns are split into two columns each. The aldehyde, fusel oil and rectifying columns use bubble cap trays. While the beer still, aldehyde column and fusel oil column can use live steam injection, the rectifying column should use a reboiler to avoid any impurities from steam getting into the product.

Plant Concept

This section discusses the building requirements and alternative plant sites and integration of the plant with the existing complex. Figure 8 presents a proposed ethanol plant building layout.



ETHANOL PLANT BUILDING LAYOUT

Size : 18m x 26m = 468m^2

FIGURE 8

Building Requirements

Table 2 summarizes the floor space and head room requirements for various plant systems. Fermentation requires the largest floor area equal to 252 sq.m. and approximately 13 m of head room. This head room is required to accommodate the five fermenters and the beer well which are approximately 10 m tall. Wet waste storage and processing area requires 80 sq.m. of space.

Distillation requires 54 sq.m. To limit the head room requirements to 18 m, two distillation columns are used. For safety reasons, this area is provided with explosion resistant walls up to the 13 m elevation.

As the remaining equipment requires lower head room, a two storey building may be used to accommodate these and service, personnel and control areas.

Total plant area requirements are approximately 514 sq.m. of floor space. By using two stories as shown in Figure 8, total main floor area requirements are reduced to approximately 468 sq.m. Therefore, a building approximately 18 x 26 m size is required with a height of 13 m except over the distillation area which is 18 m high.

Site

We have examined several potential sites for the facility. Our first inclination was to reuse the place presently occupied by Raw Potato Storage Building Number 3 (RPS 3) since we understand this building is scheduled for demolition. Problems with this site include:

1. Adjacent buildings (RPS Nos. 2 and 4) could at best be assumed to be designed for snow drifts caused by RPS No. 3, which is approximately 5 m high. The ethanol plant requires approximately 13 m over the fermentation area and 18 m around the distillation columns, and would, therefore, impose snow loads in excess of design capacity on these structures. Options would include reinforcing of the existing structures, allowance for a gap between new and existing, or use of a lower roof level on each side of the new plant (offices, storage, etc.).

TABLE 2

FLOOR SPACE AND HEADROOM REQUIREMENTS

	Length	x	Width	=	Area	Headroom
	m		m		m^2	m
Fermentation					252	13
Distillation	9		6		54	18
Control Room	6		3		18	6
Office	3		3		9	3
Change, Washrooms	6		5		30	3
Electrical Room	3		6		18	6
Stairs (2)	3		2 x 2		12	13
Truck Bay	4		8		32	6
Wet Waste Storage	8		10		80	13
Chemical Storage	3		3		9	6
					514	

Note: This estimate is for the fuel grade base case scenario.

Total Main Floor Area = 18 m x 26 m = 468 m^2

2. The existing foundation system and slab-on-grade will be similar to what was found beneath RPS No. 1. The new plant will require competent support not only for the structure but for numerous tanks and heavy equipment. Settlement criteria dictates footings on the Carberry sand (approximately 1 1/2 m below grade) which would eliminate any consideration of retention of existing concrete.

3. The space occupied by RPS No. 3 has an area of 3,530 sq.m., which is much more than required. Even if a 9 m wide corridor between RPS Nos. 2 and 4 is excluded, the area is still 3,122 sq.m. Placing a 468 sq.m. ethanol plant in this location would probably waste some of the remaining area.

4. Locating the alcohol plant here would preclude the future reuse of this space for potato storage requirements. Access to the raw potato flume would be a big advantage if additional potato storage becomes necessary.

These problems with this site indicate that it is more economical to site the plant elsewhere. Potential sites should be judged for proximity to:

- the production building
- the boiler
- underground services (well water, sewage lines)
- wet solid and freezer wastes

Possible options include:

- south of RPS #1 and #2

- north of the dry packaging building

- between the production and packaging buildings, especially if the old freezer tunnel is to be removed.

A final decision regarding plant site can be made during the predesign stage of this project.

Utilities and Energy Requirements

The alcohol plant would require the following utilities:

- Steam
- Electricity
- Cooling Water

Table 3 presents a breakdown of these loads. Discussions of each item follow.

Steam

Steam is used for cooking, distillation and dehydration. Respective energy loads are presented in Table 3. Including miscellaneous uses such as sterilization, total steam usage is expected to be approximately 4,500 kg/hour.

Carnation has advised us that when both production lines are operating, the main boiler is loaded to nearly full capacity. Winter heating loads are expected to use most of the second boiler's capacity. Therefore, a new boiler would be required. To allow for peak loads during start-ups, and for future use, a boiler with 9,000 kg/hr capacity should be installed.

Heat pump powered distillation and MVR distillation avoid this expense as these distillation schemes do not require a new boiler. It was assumed that the existing boiler capacity is adequate for supplying steam to the cooking and dehydration processes.

Electricity

Electricity is required for the various pumps and agitators. It is also required for:

- the hammer mill
- the reverse osmosis system
- lighting
- instrument air compressor

TABLE 3

ENERGY BALANCE

Steam Loads	Pressure (kPa)	Quantity GJ/d	MJ/L Product
Cooking	300	54.7	2.18
Distillation	300	132.2	5.26
Dehydration	700	5.0	0.20
		191.9	7.64

Electrical Loads (estimated)

Pumps	135 kW
Hammermill	37
Agitators	45
Lights	20
Compressors	37
Miscellaneous	140

414 kW x 24 hr = 9,936 kWh/d

round to 10,000 kWh/d

peak load $\frac{414}{.9}$ = 460 kVA

Cooling Loads

	GJ/d
Distillation Condensers	130
Mash Cooling	40
Fermenter Cooling Jackets	26
Dehydration	5
	201

Daily electrical usage is estimated at approximately 10,000 kWh/day and the peak demand is estimated at approximately 500 kVA.

Cooling Water

Well water at Carberry is available at 5 - 7°C for cooling. Very little water is required for the alcohol process, as the wastes are received at the desired moisture content for fermentation.

The main cooling loads are:

- distillation condensers
- mash cooling
- fermenter cooling jackets
- dehydration

Based on steam consumption, well water flow requirements are estimated to be 1,360 L/min. After use for cooling, this water is returned to the potato processing operation for reuse as preheated water and should not add to the total water requirements of the potato processing complex.

Operating Schedules and Labour Requirements

Carnation has indicated that it plans to operate the french fry plant 24 hours a day, 5 days a week during 1988, resulting in a total of 242 days of operation. A five day operation is made possible by the recently added capacity, and will save on the higher weekend labour costs. Therefore, it is likely that the alcohol plant would also operate for only 5 days a week, 24 hours a day. In reality, the plant will operate seven days a week, with staggered shutdown of fermentation and distillation. We have assumed a higher plant capacity than would result from a 5-day per week, 48 weeks per year operation. It is likely that as the market demand picks up, longer than five days per week potato plant operation would result.

Two operators would be required to operate the alcohol plant. One would operate raw material collection, cooking and fermentation. The second would be responsible for distillation, product denaturing and storage. It is anticipated that four crews would be required for a 24 hour operation. In addition, one day-shift worker would help out with tasks that can be scheduled such as cleaning, start-ups, etc. A plant manager would supervise operations.

Additional labour can be supplied by the existing potato processing operations for the following services:

- purchasing
- receiving
- sales
- secretarial
- maintenance
- stationary engineer

This reduces operating and capital costs as compared to a stand-alone ethanol plant. Expense of maintenance shops, boiler room, and a head office are avoided by integrating the ethanol plant with the potato processing facility.

Lost Value of Solid Wastes

By producing ethanol from the wet solid and freezer wastes, Carnation would lose the value of these wastes to the feed lot. However, this is expected to be small compared to the other operating costs. In reality, the cost of transporting these materials to the feed lot has often been higher than their feed value due to relatively small numbers of heads of cattle. Recently, Carnation has arranged with a contractor for hauling the wastes on the basis that the contractor will supply the required quantity of these wastes to the feedlot at $1/t and keep the remainder, thus reducing hauling costs to Carnation. The ethanol plant will offset some of the material being diverted away from the feedlot with a high protein, thick stillage by-product.

Plant Reliability

An ethanol plant is a reliable method of disposing wet solid and freezer wastes, and a back-up disposal method is not required. Surge capacity of various tanks is sufficient to allow for short shutdown for emergency maintenance.

Supplementary Feedstocks

Fermenting supplementary feedstocks in addition to the potato wastes would increase production. This could be achieved either through an increase in the plant capacity or through an increase in the number of operating hours.

Two feedstocks were considered. Starch from a potato processing plant in Winnipeg might be available at low or no cost. Molasses with a fermentable sugar content of 50 percent is available as a feedstock to the plant. Depending on the season, it should sell for as little as $85/tonne and should be considered as a supplementary feedstock. Molasses contains approximately 54 percent fermentable sugars, mostly sucrose.

Two scenarios were considered. The first was to ferment molasses on days when the french fry plant is shutdown. This permits more continuous operation and also produces more product.

The second scenario is to ship the starch from Winnipeg to Carberry once or twice a week and to ferment it without increasing operating hours. This would increase throughput and production by approximately 9 percent without increased labour costs and with a small increase in capital costs. Profitability of these two scenarios was compared with the base case, and the results are presented elsewhere in the report.

TABLE 4

CAPITAL COST ESTIMATE FOR FUEL GRADE BASE CASE

Item	Capital Cost 1,000 $ Cdn., 1987
Reverse osmosis system	$ 150
Hydrocyclone, clarifier system	100
Cooking, fermentation, product storage	1,480
Distillation system	1,400
Building	500
Utilities and raw material transfer	500
SUB-TOTAL	4,130
Construction indirects	413
Engineering	413
	$4,956
TOTAL	$4,956,000

TABLE 5

ANNUAL OPERATING COST ESTIMATE
FOR FUEL GRADE BASE CASE

	$1,000 Canadian 1987
Labour	
9 workers x $30,000/year	$ 270
Alcohol plant manager $45,000/year	45
Shared labour costs (purchasing, receiving, sales, secretarial, maintenance, stationary engineer)	100
Utilities	
Steam	280
Electricity	85
Molecular sieve adsorbent replacement	45
R.O. membrane replacement	11
Enzymes and chemicals	200
Maintenance materials (2.0 percent of capital costs)	100
Feedstock costs	-
	$1,136
TOTAL	$1,136,000

COST ESTIMATES

Capital and operating costs have been estimated for the following two products:

- Fuel Alcohol
- Food/Industrial Alcohol

In each case, a number of scenarios have been formulated which are a variation of the base cases presented in the previous section.

Fuel Alcohol

Table 4 presents the capital cost estimate for producing fuel grade alcohol. This is the base case using steam for distillation and no supplementary feedstocks. The total cost is estimated at $4,956,000 and includes:

- equipment
- installation materials and labour
- taxes
- building
- engineering
- construction indirects (contractor costs and profit)

A 25 percent contingency allowance is included in each separate entry as the estimated accuracy is expected to be ±25 percent.

Table 5 presents an estimate of annual operating costs for producing anhydrous fuel grade alcohol assuming Scenario 1, the base case. Labour costs were calculated based on an average labour cost of $30,000 including benefits and overhead. Steam costs are calculated based on the current natural gas cost of $9.7/cu.m. and a boiler system efficiency of 65 percent. This low efficiency was assumed because most condensate would not be returned to the boiler. Both cooking and distillation use live steam injection.

Capital and operating cost estimates were also prepared for several other fuel grade ethanol scenarios and are summarized in Table 6. The Base Case is Scenario 1. Scenario 2 is for purchasing molasses to extend production to 335 days per year. This would require little extra capital costs. However, the operating costs are substantially higher than the Base Case. Based on a price of $85/tonne and a fermentable sugar content of 54 percent, annual molasses costs would be $381,000. Two additional workers are required to operate the plant for 335 days per year. Additional costs for steam, electricity, and increased maintenance are also included in the operating costs.

Scenario 3 is for using MVR powered distillation to produce fuel grade alcohol. Higher equipment costs for a compressor and a low pressure drop distillation column are partially offset by reduced utilities cost as an additional boiler is not required. This scenario has lower operating costs than the Base Case because MVR distillation is used. The high COP of the MVR system replaces the steam cost for distillation with a smaller electrical cost. Some steam would be required because the dehydration system requires condensing the product in a separate condenser and compression and recovery of heat contained in this product stream is not likely to be viable.

Scenario 4 uses a high temperature heat pump to power fuel grade distillation. Similar to MVR, the cost of a heat pump is partially offset by the reduced utility systems costs. It also has lower operating costs than Scenario 1 because of the higher energy efficiency of the heat pump powered distillation. However, the operating costs are a bit higher than for MVR because of the lower COP's.

Scenario 5 uses batch fermentation. Equipment costs are lower as mild steel fermenters can be used. However, more fermenters require more processing room, resulting in a higher building cost. Batch fermentation also has higher peak steam loads, thereby higher utility system costs are incurred. Total capital costs are also slightly higher than the Base Case. This scenario would require one additional worker each shift for cleaning fermenters and assisting the fermentation operator. Costs for higher maintenance, energy usage and cleaning chemicals would also increase the operating costs for the batch fermentation case.

TABLE 6

CAPITAL AND OPERATING COSTS AND REVENUES
FOR FUEL GRADE ALCOHOL PRODUCTION SCENARIOS

	Fuel Base Case	Fuel 335 days/yr (Molasses)	Fuel MVR	Fuel Heat Pump	Fuel Batch Fermentation	Fuel 280 days/yr (Old Dutch Starch)	Fuel Conventional Waste Pretreatment	Fuel Peel Waste Impact
Scenario	1	2	3	4	5	6	7	8
Capital Cost ($1,000 Cdn 1987)								
Direct								
Building and Site Prep	$ 500	$ 500	$ 500	$ 500	$ 650	$ 500	$ 500	$ 600
Equipment	3,130	3,230	4,480	6,010	3,069	3,230	2,960	3,675
Utilities, etc.	500	500	375	400	600	500	500	600
TOTAL DIRECT COSTS	4,130	4,230	5,355	6,910	4,319	4,230	3,960	4,875
Indirects								
Construction	413	423	535	691	432	423	395	487
Engineering	413	423	535	691	432	423	395	488
TOTAL INDIRECT COSTS	826	846	1,070	1,382	864	846	790	975
TOTAL CAPITAL COST	$4,956	$5,076	$6,425	$8,292	$5,183	$5,076	$4,750	$5,850
Operating Costs ($1,000 Cdn 1987)								
Raw Material	$ -	$ 381	$ -	$ -	$ -	$ -	$ -	$ -
Other	1,136	1,288	992	1,037	1,300	1,249	1,120	1,322
TOTAL OPERATING COSTS	$1,136	$1,669	$ 992	$1,037	$1,300	$1,249	$1,120	$1,322
Revenue								
Production (L/y) / 1,000	7,000	8,375	7,000	7,000	7,000	7,600	6,200	8,770
Base Price	$.30/L	$.30/L	$.30/L	$.30/L	$.30/L	$.30/L	$.30/L	$.30/L
Revenue ($/y)	2,100	2,512	2,100	2,100	2,100	2,280	1,860	2,631

Scenario 6 involves shipping waste starch from Winnipeg to Carberry for use as a supplementary feedstock. Capital costs are slightly higher than for Scenario 1, to allow for shipping containers and material handling equipment. However, additional operating costs are required for increased utility consumption, increased chemical consumption and shipping costs. Trans-X was contacted for an estimate for shipping approximately 40,200 kg of starch slurry weekly and returning the empty containers to Winnipeg. The cost is estimated at approximately $1,200 a week, or $62,000/year. Increases in operating costs totalled $113,200, which amounts to only $.19/L in increased production over the 7 million litre per year base case. This low cost occurs mostly because of no additional labour or maintenance costs.

Scenario 7 involves the use of only the conventional approach to collecting and processing wastes. That is, membrane technology is not used for preconcentrating blanching and second peel scrubber streams. As the low concentration of these streams will result in low ethanol concentration, thereby high capital cost and high energy costs, it is assumed that these streams will not be used. However, the first peel scrubber effluent, which is at a higher concentration of 1.4 percent can be used while keeping the sugar concentrations to within acceptable limits. Some reduction in capital costs, will occur because of the elimination of the membrane system and reduced production capacity. Similarly, operating costs will also decrease because of the above reasons. The key negative factors include approximately 14 percent reduction in production capacity, and an increase in total organic load to the lagoons.

As discussed earlier, testing program carried out during this study could not accurately establish the total qualities of peel wastes. Carnation estimates their peel losses at 11 percent of feed. Assuming that these losses occur through peel sludge, first peel scrubber and second peel scrubber streams, a substantial quantity of carbohydrates remains unaccounted for. Scenario 8 estimates this quantity and assumes that it is available to the ethanol plant. Result is a higher capital cost, a higher operating cost, but a substantially higher production rate than the base case.

Food/Industrial Ethanol

Table 7 presents capital and operating costs and revenues for four scenarios for food/industrial ethanol production.

Scenarios 9 and 10 use the same assumptions regarding waste pretreatment as the fuel alcohol Base Case. For Scenario 9, an azeotropic product is produced. The distillation process shown in Figure 7, is used, and has formed the basis of capital cost estimation. Although distillation is more complex than the fuel ethanol base case, no dehydration system is required. Despite this, total capital costs are higher than the fuel Base Case, as are the operating costs because of higher energy and labour costs.

Scenario 10 which assumes production of anhydrous alcohol, has higher capital and operating costs than Scenario 9 as it needs a dehydration system. This scenario has been included as pure anhydrous ethanol is required in some applications. The higher price for this special product might make the additional costs worthwhile.

Scenario 11 makes the same assumptions regarding waste pretreatment and use as Scenario 7. Thus, very dilute streams are not used and a membrane concentration system is avoided. It produces azeotropic food/industrial grade product. Capacity is lower than Scenario 9, as are capital and operating costs. Scenario 12 is similar to Scenario 8 in that all peel wastes are utilized. An azeotropic food/industrial grade product is produced. Capital and operating costs and revenues are higher than those for Scenario 9 because of the higher production capacity.

TABLE 7

CAPITAL AND OPERATING COSTS AND REVENUES
FOR FOOD GRADE ALCOHOL PRODUCTION SCENARIOS

	Food/Industrial Base Case (Azeotropic)	Food/Industrial (Anhydrous)	Fuel Conventional Waste Pretreatment	Fuel Peel Waste Impact
Scenario	9	10	11	12
Capital Cost ($1,000 Cdn 1987)				
Direct				
Building and Site Preparation	$ 600	$ 600	$ 600	$ 700
Equipment	3,575	3,900	3,360	4,220
Utilities, etc.	550	550	550	650
TOTAL DIRECT COSTS	4,725	5,050	4,510	5,570
Indirects				
Construction	473	505	450	555
Engineering	473	505	450	555
TOTAL INDIRECT COSTS	946	1,010	900	1,110
TOTAL CAPITAL COST	$5,671	$6,060	5,410	6,680
Operating Costs ($1,000 Cdn 1987)				
Raw Material	$ -	$ -	$ -	$ -
Other	1,277	1,342	1,258	1,468
TOTAL OPERATING COSTS	$1,277	$1,342	$1,258	$1,468
Revenue				
Production (L/y)	7,370,000	7,000,000	6,530,000	9,230,000
Base Price	$.35/L	$.40/L	$.30/L	$.30/L
Revenue ($/y)	2,580	2,800	1,959	2,769

FINANCIAL ANALYSIS

Wardrop's cashflow model was used to carry out a financial analysis of the various fuel and food-grade ethanol scenarios presented in Tables 6 and 7 respectively. Results of these analyses are presented in this section under the following headings:

- Basic Assumptions
- Results
- Sensitivity Analyses
- Conclusion

Appendix B presents the methodology used to develop the Wardrop's cashflow model, as well as the results of the various sensitivity analyses conducted.

Basic Assumptions

The following assumptions and input data have been used:

- 1 year construction period
- 20 year project life
- Zero percent inflation rate
- 15 percent CCA on equipment, 5 percent on buildings, declining balance
- 2:1 debt to equity ratio
- 4 percent (inflation free) interest rate
- Working capital equal to three months of operating cost
- Start-up expenses of $150,000
- 50 percent corporate tax rate
- Project to be undertaken by an existing corporation with appropriate tax consequences

Capital and operating costs and production capacities for various scenarios were used as presented in Tables 6 and 7. After-tax IRR calculations were made over a range of prices, and results are presented in plots of IRR versus ethanol prices.

Table 8 presents the required selling prices for the various scenarios at 30 percent after-tax IRR for comparison purposes.

TABLE 8
REQUIRED SELLING PRICES FOR 30 PERCENT AFTER-TAX IRR

Scenario	Title	Price $Cdn/L
Fuel Grade Ethanol		
1	Base Case	0.30
2	Molasses	0.33
3	MVR	0.32
4	Heat Pump	0.37
5	Batch Fermentation	0.34
6	Old Dutch Starch	0.30
7	Conventional Waste Pretreatment	0.34
8	Peel Waste Impact	0.28
Food Grade Ethanol		
9	Base Case Azeotropic	0.33
10	Anhydrous	0.36
11	Conventional Waste Pretreatment	0.36
12	Peel Waste Impact	0.30

After-Tax IRR - Fuel Ethanol

After-tax IRR's for the various fuel ethanol scenarios are presented in Figure 9A and 9B. The Base case which involves steam-driven distillation and continuous fermentation, waste starch supplementation, and peel waste utilization are the three most attractive options. An FOB selling price of $.30/L will result in after tax IRR's of 28 to 34 percent for these three cases. The MVR option is also comparable to the base case in the lower price range.

Heat pump, molasses supplementation and batch fermentation scenarios will require a $.35/L price to be in the 25 to 30 percent IRR range. Key reasons are:

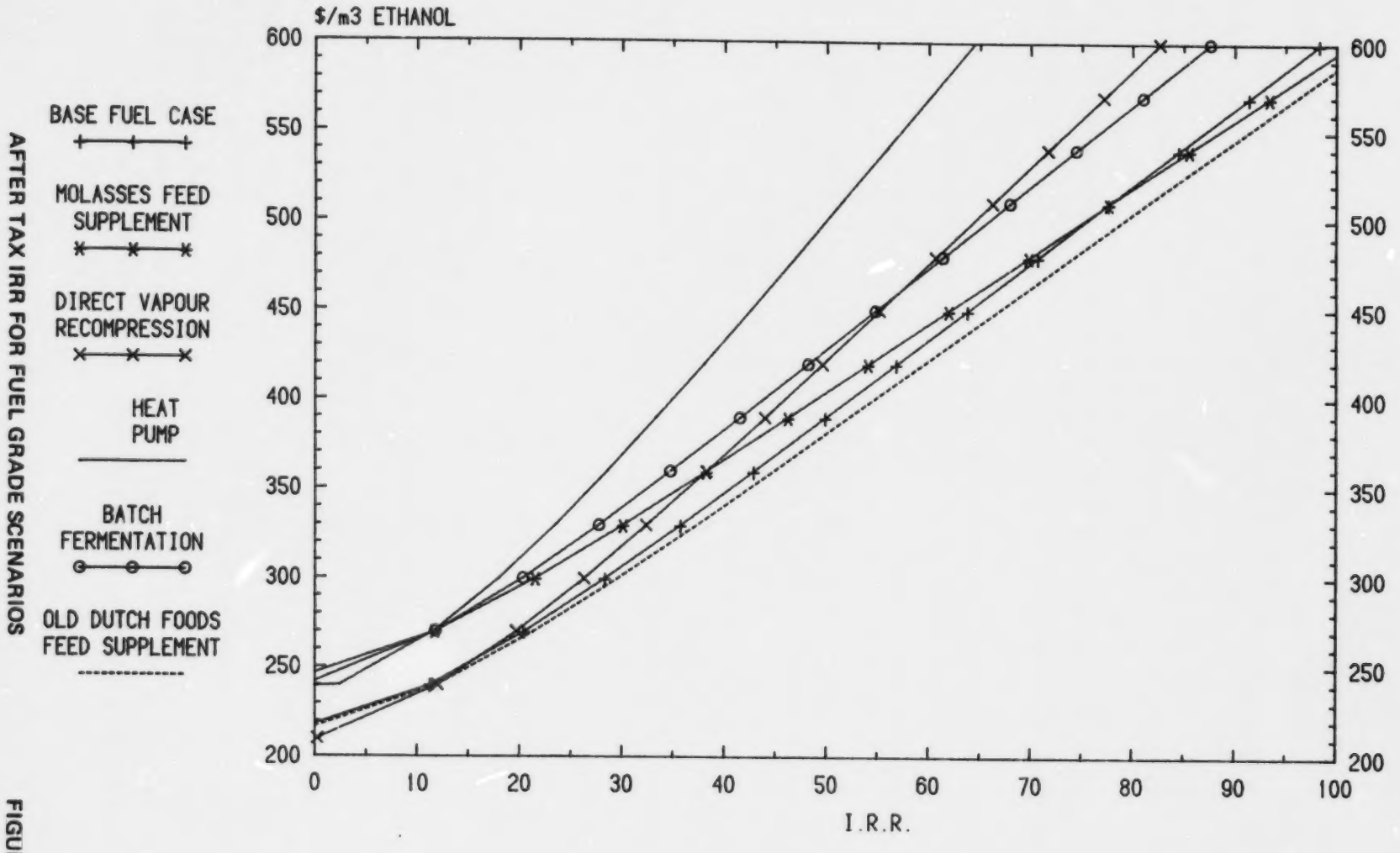
- Higher capital cost for the heat pump.
- Higher capital and labour costs for batch fermentation.
- Relatively high cost of molasses.

Scenario 7, which assumes not using reverse osmosis to concentrate blancher and second scrubber wastes, requires a selling price of $.34/L to obtain an IRR of 30 percent. This is because these dilute wastes are discarded resulting in 11 percent less production than the base case.

Thus, it would appear that an FOB price range of $.30 to $.35/L can result in a viable fuel ethanol project.

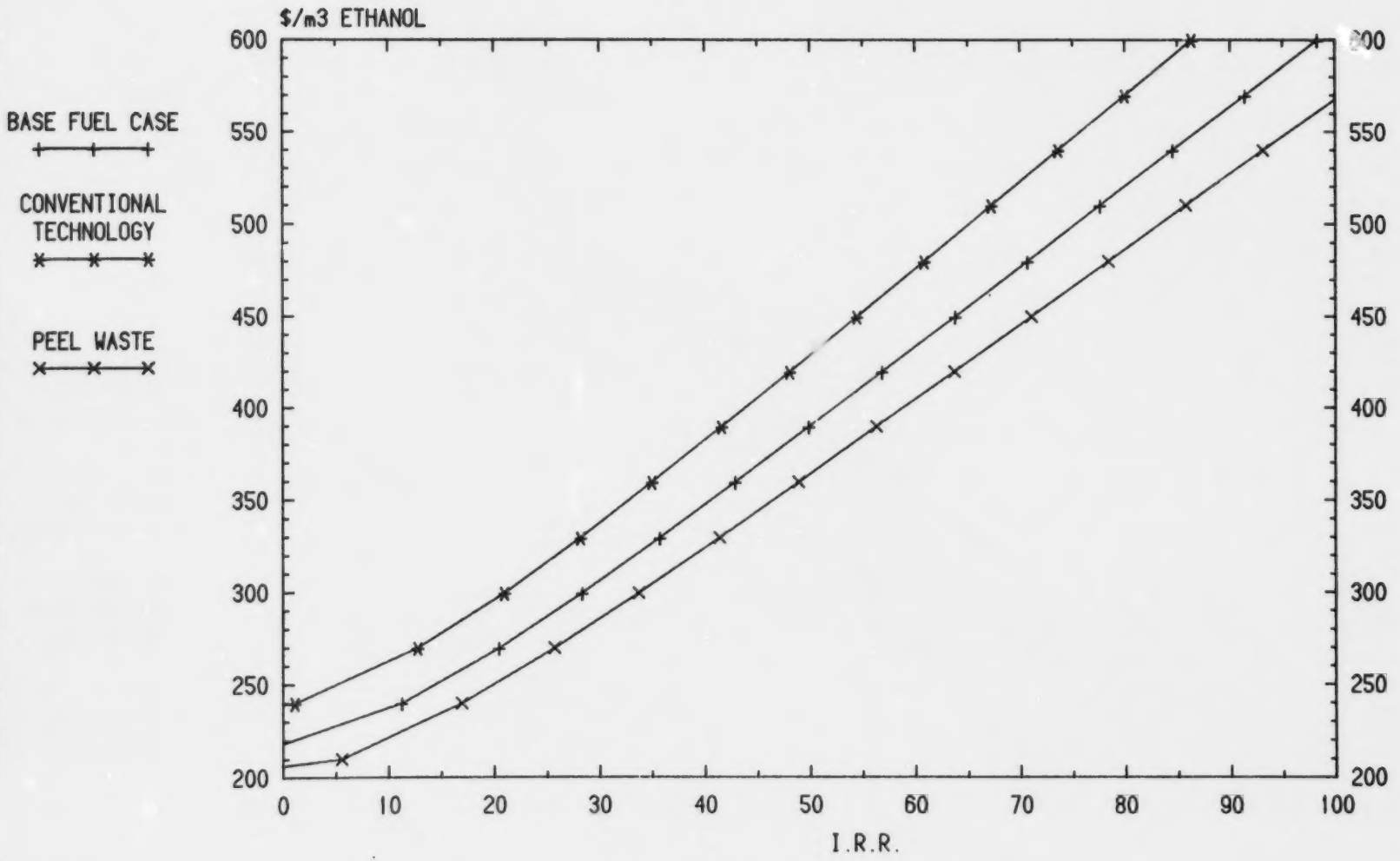
After-Tax IRR - Food/Industrial Ethanol

Figure 10 presents the results for the four food/industrial grade ethanol scenarios. Each involves continuous fermentation with steam-driven distillation. After-tax IRR's in the range of 25 to 40 percent are obtained at a FOB plant price of $.35/L. The base food grade case requires a selling price of $.33/L for an IRR of 30 percent. The anhydrous scenario requires a higher selling price, $.36/L, because of the higher capital and operating costs. Peel waste utilization results in the most attractive IRR and the lowest required selling price $.30/L because of increased production. Not using reverse osmosis and discarding the dilute wastes requires a selling price $.03/L higher than the base case, due to lower production.



AFTER TAX IRR FOR FUEL GRADE SCENARIOS

FIGURE 9A



AFTER TAX IRR FOR ADDITIONAL FUEL GRADE ETHANOL SCENARIOS

FIGURE 9B

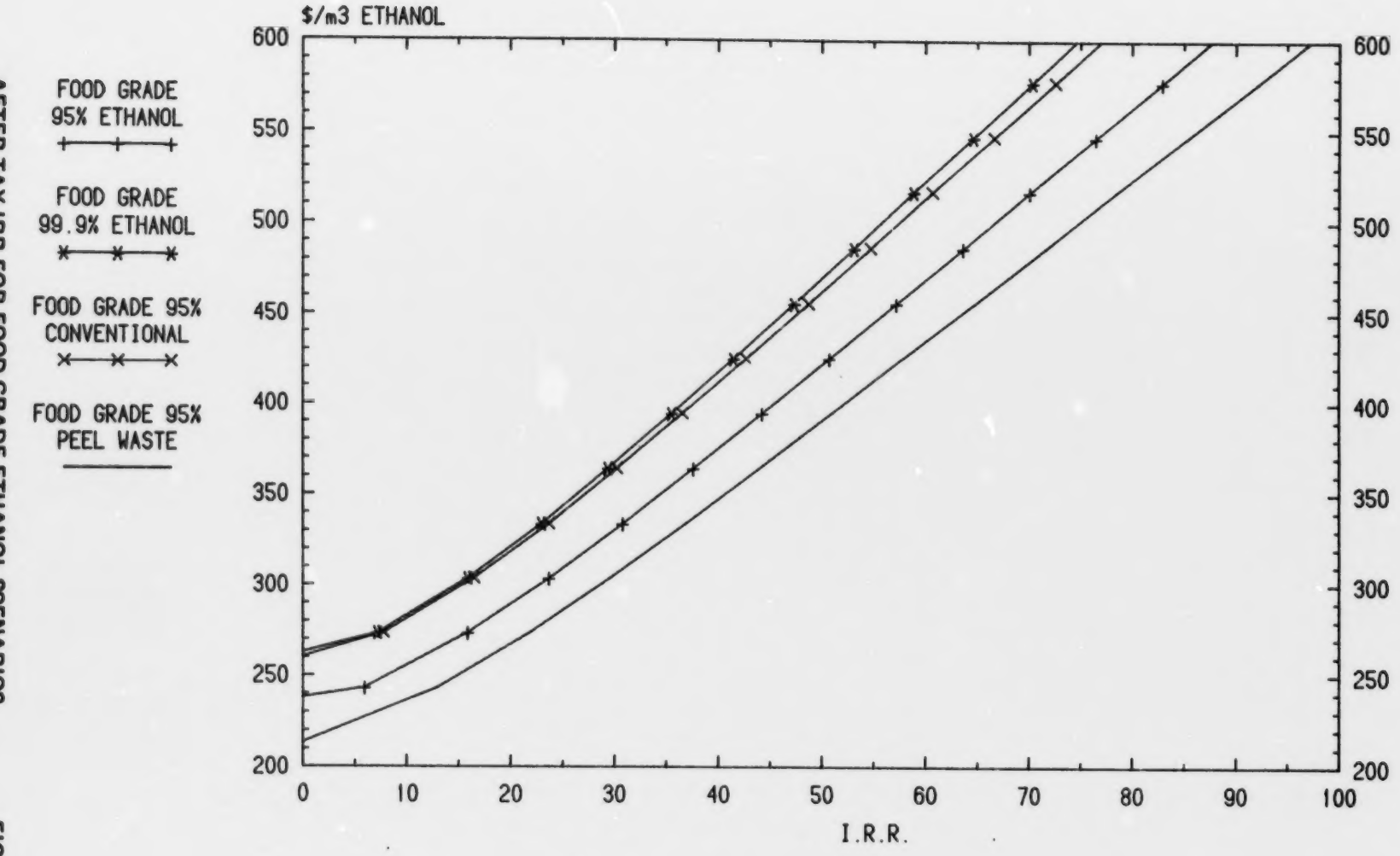


FIGURE 10 AFTER TAX IRR FOR FOOD GRADE ETHANOL SCENARIOS

If industrial ethanol is the preferred product, a combination of the two products, i.e. bulk of the production being azeotropic ethanol, and small amount of anhydrous ethanol, may be desirable to improve marketing flexibility.

Although fewer variations of the base case scenario have been evaluated for food/industrial ethanol, the effects can be estimated by comparison with the same analysis for the fuel ethanol case. Thus, for example, it is expected that the IRR will improve if Old Dutch Foods' starch is used, and will reduce with batch fermentation.

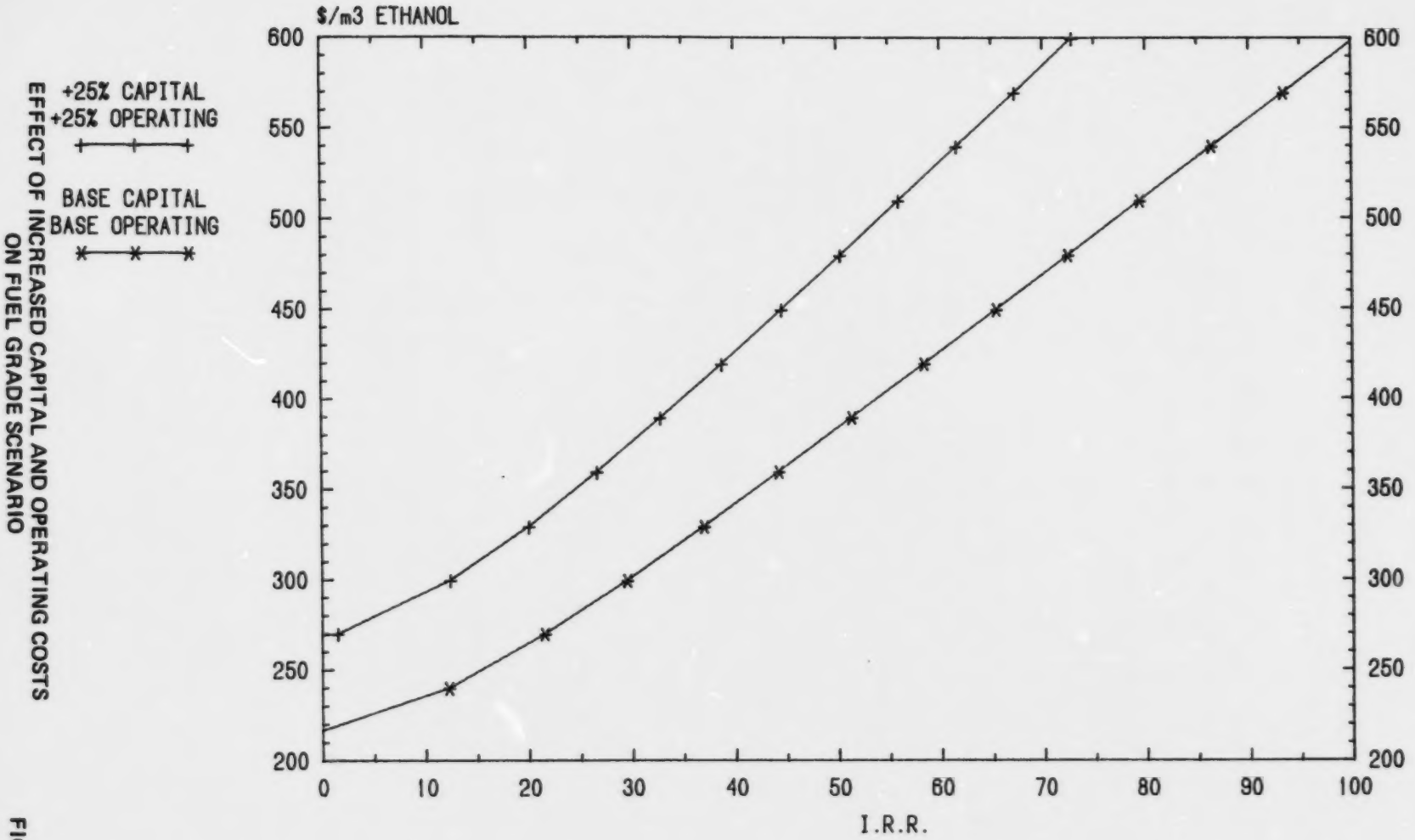
Sensitivity Analyses

Sensitivity analyses have been performed for both fuel and food/industrial base case scenario for the following parameters:

- Capital cost (±25 percent)
- Operating cost (±25 percent)
- Production capacity (±10 percent)
- Interest rate (10 percent, zero inflation)
- Debt to equity ratio (1:1)

Results are presented in Appendix B in the form of after tax IRR versus ethanol price charts. Parameters with the largest impact appear to be operating costs and debt to equity ratio.

One sensitivity analysis, carried out to assist in risk analysis, involves a 25 percent higher capital and operating costs. This is presented in Figures 11 and 12 for fuel and food/industrial cases respectively. It would appear that in both cases, approximately $.10/L product price increase will be required if a 25 - 30 percent after tax rates of returns are desired.



EFFECT OF INCREASED CAPITAL AND OPERATING COSTS ON FUEL GRADE SCENARIO

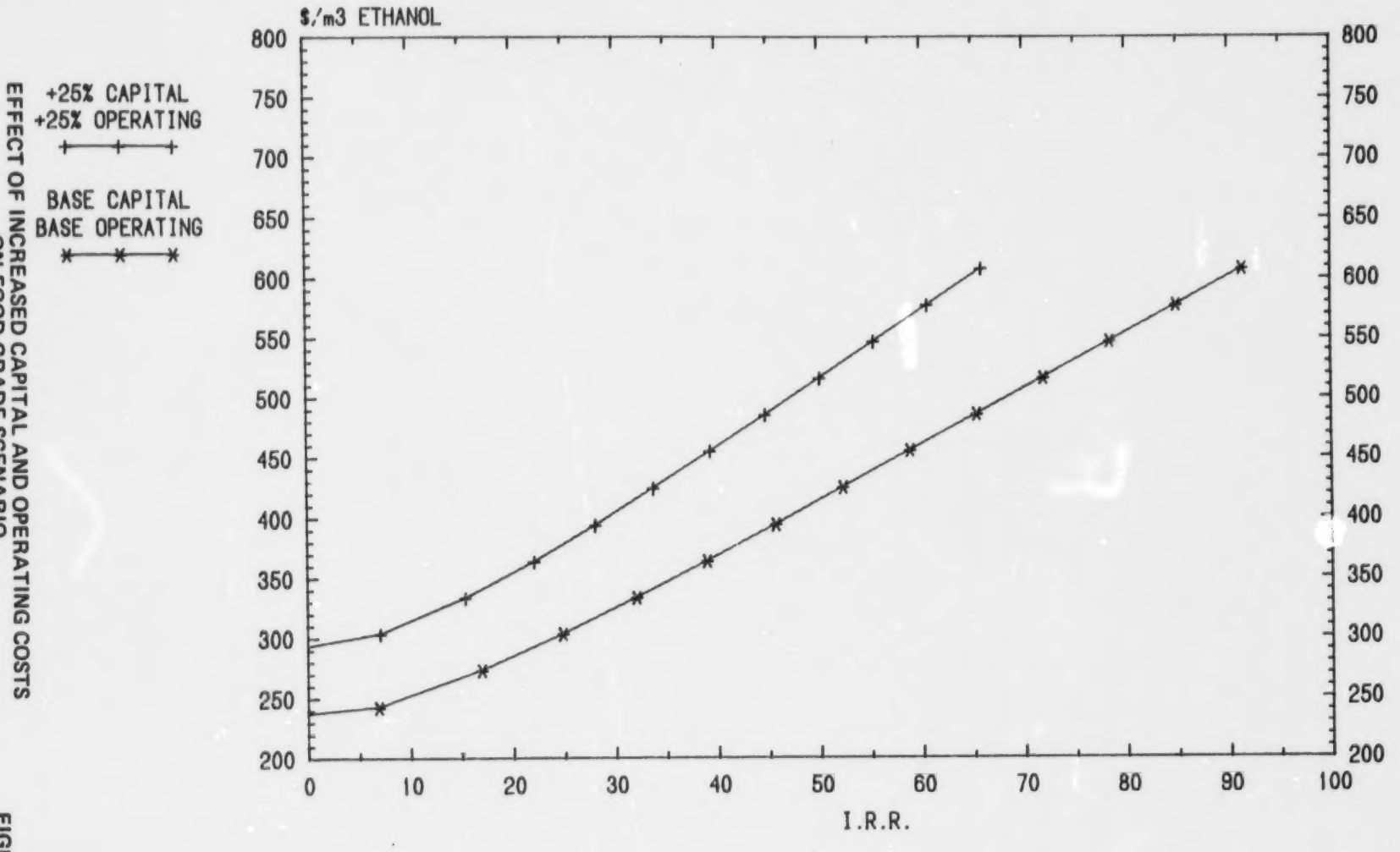
FIGURE 11

Conclusions

Our financial analyses indicate that 28 and 34 percent after-tax internal rates of returns can be obtained at FOB plant prices of $.30 and $.35/L for fuel and food/industrial grade products respectively. A $.30/L fuel grade ethanol price is attractive for methanol-ethanol-gasoline blending at current gasoline prices. Similarly, $.35/L price for food/industrial grade ethanol is competitive for bulk U.S. prices for this commodity, and lower than the present Canadian price.

A risk analysis of this project should consider a maximum price of $.40/L for both fuel grade and for food/industrial grade products. Through suitable negotiations with fuel blenders and vinegar producers, Carnation should be able to obtain these prices. However, further project development efforts should be aimed at controlling factors that will lead to these higher prices.

Based on this analysis, it is our opinion that production of ethanol from wastes at Carnation's plant has a very high viability potential.



EFFECT OF INCREASED CAPITAL AND OPERATING COSTS ON FOOD GRADE SCENARIO

FIGURE 12

PROJECT IMPLEMENTATION

This analysis indicates that an excellent potential for viability exists for fuel or food/industrial grade ethanol production at Carberry. However, a number of technical uncertainties must be resolved. Examples are:

- Reverse Osmosis System: The RO system design is based on bench-scale data obtained at CFPDC. Analysis using Wardrop's cost optimization model indicates that Filmtec spiral wound membranes result in the most cost-effective system. On-site pilot-scale tests over an extended period (perhaps 3 months) are required to confirm the performance of these membranes. Permeate recycle should also be evaluated during these tests.

- Because of the difficulties in sampling, we have neglected peel sludge and first scrubber wastes from our yield calculations. Published data, and our monitoring experience at a similar facility indicate that substantial quantities of carbohydrates are being rejected in these streams. An analysis of a scenario using peel wastes indicates that improved viability results. More detailed monitoring and fermentation work is required during the next stage of the ethanol project to establish quantities of peel wastes.

- Ethanol yields have been calculated using conventional data. However, actual fermentation tests are required to establish yields more accurately.

- Effect of the ethanol plant on the wastewater loading to the lagoons should be estimated through experimental work.

Most of these issues will be addressed by CFPDC and Wardrop as part of a Bioenergy Development Program sponsored project which has started in the fall of 1987. This includes the design and construction of a membrane system pilot plant and on-site testing with the various streams of interest. Fermentation work is also planned as part of this project on the various waste streams.

This will help define the quantities of carbohydrates in the peel wastes, and yields of ethanol from the various streams. The fermentation tests will also help define the wastewater characteristics.

Key questions that affect capacity include the peel waste quantities and plant operating schedule. As discussed earlier, peel waste could add approximately 25 percent to the ethanol plant capacity and its use will improve plant viability. During the last few years, the potato plant has operated for 280 days per year or more. Even after doubling of the capacity during the fall of 1987, seven days per week operation was implemented till Christmas. However, projection for the 1988 year is 242 days of operation. The operating days in the following years will very much depend on the rate at which product markets are developed. If past history is an indication, 280 days per year operation could result. We have evaluated addition of waste starch from Winnipeg and molasses to increase capacity and to increase the number of ethanol plant operating days. The first factor enhances viability, while the second depends on the cost of molasses. These issues should be reviewed periodically as the project develops.

A number of companies have showed an interest in participating in the project and/or buying the product. A key issue is whether the plant will produce fuel or food-grade ethanol. As Carnation is not involved in the fuel or chemical business, it does not wish to own and operate the ethanol plant. However, it is interested in negotiating a suitable arrangement with a company to lead to the production of ethanol at its site.

Thus, Carnation should pursue the following course of action:

- Fully cooperate with the BDP sponsored project to determine the applicability of the membrane system, to evaluate permeate recycle, and to determine yields and potential plant capacity.

- Invite proposals from interested companies regarding the production of ethanol at their Carberry site using potato plant waste streams.

- Revise the technical and economic evaluation following the completion of the above two activities to confirm the results of this work, and to develop the most viable plant concept from the standpoint of wastes utilized, operating schedule, feedstock supplementation and integration of the plant with the existing complex.

A proposed target date for starting the preliminary design of the plant is October 1, 1988, as it should be possible to complete these activities by that date. If this target is met, the ethanol plant commissioning can start on April 1, 1990. Full production can then resume with the new potato season in September 1990.

CONCLUSIONS

We conclude that:

1. An estimated seven million litres of ethanol can be produced at Carnation Foods Carberry Plant annually using the various liquid and solid waste streams as feedstock.

2. Markets exist for both fuel and food/industrial grade products, and a number of companies are interested in discussing participation in the project and/or buying the product.

3. A number of technical uncertainties exist. These can be resolved through the Bioenergy Development Program sponsored project which is now underway.

4. The base case fuel scenario results in an after-tax internal rate of return of 28 percent at $0.30 per litre selling price.

5. After-tax internal rates of return in excess of 30 percent can be obtained for the food grade product at a price of $0.35 per litre.

6. Supplementation of plant raw material with starch from a Winnipeg potato processing plant enhances project viability. Supplementation with molasses does not.

7. Use of mechanical vapour recompression or heat pump to replace steam for distillation is not economical at the low cost of natural gas, recently negotiated by Carnation.

8. If both capital and operating costs increase by 25 percent, an increase in ethanol price of $.10/L will be required to maintain an after-tax IRR of 30 percent.

9. Use of peel waste improves project viability, and quantities of these streams should be established through a detailed sampling and analysis program.

RECOMMENDATIONS

We recommend that Carnation actively pursue the production of ethanol from its waste streams. Project implementation aspects have been discussed in the report, and the following course of action is recommended.

1. Cooperate with the Bioenergy Development Program sponsored pilot scale membrane testing and fermentation project.

2. Solicit proposals from the fuel and food grade ethanol users regarding their investment in the project and marketing the product.

3. Revise the capacity and economic evaluation presented in this report following the above actions. If results remains positive, the design of the facility can be initiated by the sponsoring company.

APPENDIX A

MASS BALANCE

A. Reverse Osmosis System Mass Balance:

Feed to R.O. System:

7,300 kg/d CHO
1,780,000 kg/d total
0.4% CHO

RO Concentrate:

7,300 kg/d CHO
165,000 kg/d total
4.4% CHO

B: Total Mass into Ethanol Plant:

42,900 kg/d CHO
14% CHO
306,400 kg/d total

C: Saccharification and Fermentation should have 92% conversion efficiency.

Produce: 20,100 kg/d ethanol (7% wt (8% vol) in beer)
21,000 kg/d CO_2
1,500-4,500 kg/d yeast, non-fermentables
90 kg/d fusel oil

D: Distillation should obtain 98% recovery.

Produce: 19,700 kg/d ethanol, containing approximately 0.4% moisture.

Waste product is 260,000 kg/d stillage (mostly water, unfermented solids, yeast)

By-product is 180 kg/d fusel oil (50% F.O., 50% H_2O)

E: Product Out

19,700 kg/d, 25,100 L/d ethanol
operating 280 days a year would produce 7,000,000 L/y ethanol

LAGOON WATER BALANCE

Water discharged to lagoon - present operations:

	kg/d H_2O
Second scrubber wastes	396,800
Blanching Wastes	1,380,000
Cutting Wastes	597,600
	2,374,400

After ethanol plant built:

Stillage	267,800
Cutting Wastes	592,400
	860,200

Net reduction in water discharge to lagoon = 1,514,200 kg/d

APPENDIX B

ECONOMIC ANALYSES

This Appendix describes the economic analyses method used in the financial evaluation of the ethanol from waste starch facility proposed for Carnation's potato processing plant at Carberry, Manitoba.

Methodology

Wardrop's cashflow computer model was used to calculate cashflow tables, project net present values and internal rates of return (IRR) and for ethanol production. Several scenarios were run to determine sensitivity of IRR to various financial parameters.

Mathematical Model and Input/Output Parameters

The following data entries were used for execution of the economic analysis computer program:

1. Construction period (years)
2. Project lifetime (years, including construction period)
3. Capital item costs and associated inflation rates (per year)
4. Capital cost allowance or depreciation rates (percent per year, declining balance) for each capital item
5. Debt to equity ratio for each capital item
6. Interest rate on borrowed capital
7. Operating costs and associated inflation rates
8. Raw material costs and associated inflation rates
9. Percent of annual operating cost required for working capital and associated debt to equity ratio and interest rate
10. Product (ethanol) annual production rates, revenue per unit of product and associated inflation rate
11. Project start-up cost incurred in start-up year
12. Corporate tax rate and whether project is operated as a "stand-alone" or as a subsidiary activity

Description of entries used in this report analyses are summarized and below:

Inflation and Interest: Annual inflation rates of 0 percent and interest rates of 4 percent are used for all analyses performed in this report except for one case where 10 percent and 0 percent inflation is assumed to show sensitivity of project finances to interest rate on borrowed capital.

Capital Cost Allowance: The capital cost allowance (CCA) factor for each capital item is calculated based on its allowances in the following manner:

- The CCA factor at each year "n" for an item spent during the same year is half the allowances for that year.

- The CCA factor at year "n" for an item spent in year "j" is:

$$Dj_1n = (1-(...(1-(1-(1-0.5\,d_j)d_{j+1})d_{j+2})...)d_{n-1})\,d_n$$

where Dj_1n = CCA factor from year "j" to year "n"

d_i = CCA for year "i"

The annual total CCA amount for each item is the sum of the product of the inflated cost and the CCA factor for all previous years. A total CCA is calculated as the sum of these itemized CCA's.

The itemized capital cost debts are summed to give the total capital cost borrowed. Before beginning debt repayments, the itemized interest on these debts is calculated using the inputted interest rates. These itemized interests are summed to yield the total annual interest payments. A debt to equity ratio is applied on these interests and added to the total debt.

During the repayment period, the annual installment is calculated for each item as:

$$R = D + i \frac{(1 + i)^n}{((1 + i)^n - 1)}$$

where R = annual fixed installment, $
D = total debt, $
i = annual interest rate, percent
n = repayment period, years

The interest paid portion of this installment is the product of the unpaid portion of the debt and the interest rate. The principal repayment constitutes the balance. Total annual installments, interest and principal, are the sum of their itemized components.

Operating and raw material costs are calculated as the sums of the product of the itemized rates (eg. Table A-1).

Annual working capital is calculated as 25 percent of the sum of the operating costs and the raw material costs, corresponding to a three month reserve. Working capital is entered as total equity for all analysis in this report.

Start-up costs are incurred at the start-up year of the project, year two for all cases in this report.

Revenue from products are calculated as the sums of the product (ethanol) production rate and associated price.

Taxable annual income is the product's revenue minus the sum of :

1. The total depreciation (CCA)
2. Start-up costs
3. Repaid interest on the capital costs
4. Working capital interest (if any)

5. Operating costs
6. Raw materials costs (if any)
7. Construction interest

Taxes for a subsidiary activity are the product of the taxable income and the entered tax rate. For an independent or stand-alone activity, a cumulative taxable income may be calculated. No taxes are imposed in such cases till the year where this cumulative taxable turns positive. Taxes for an independent activity are then calculated on this cumulative value for that year and on the annual taxable income for subsequent years.

Present value for the project is calculated as:

$$P = \frac{C_n}{(1 + d)^n}$$

where: Cn is the after-tax profit for year n and d is the discount rate.

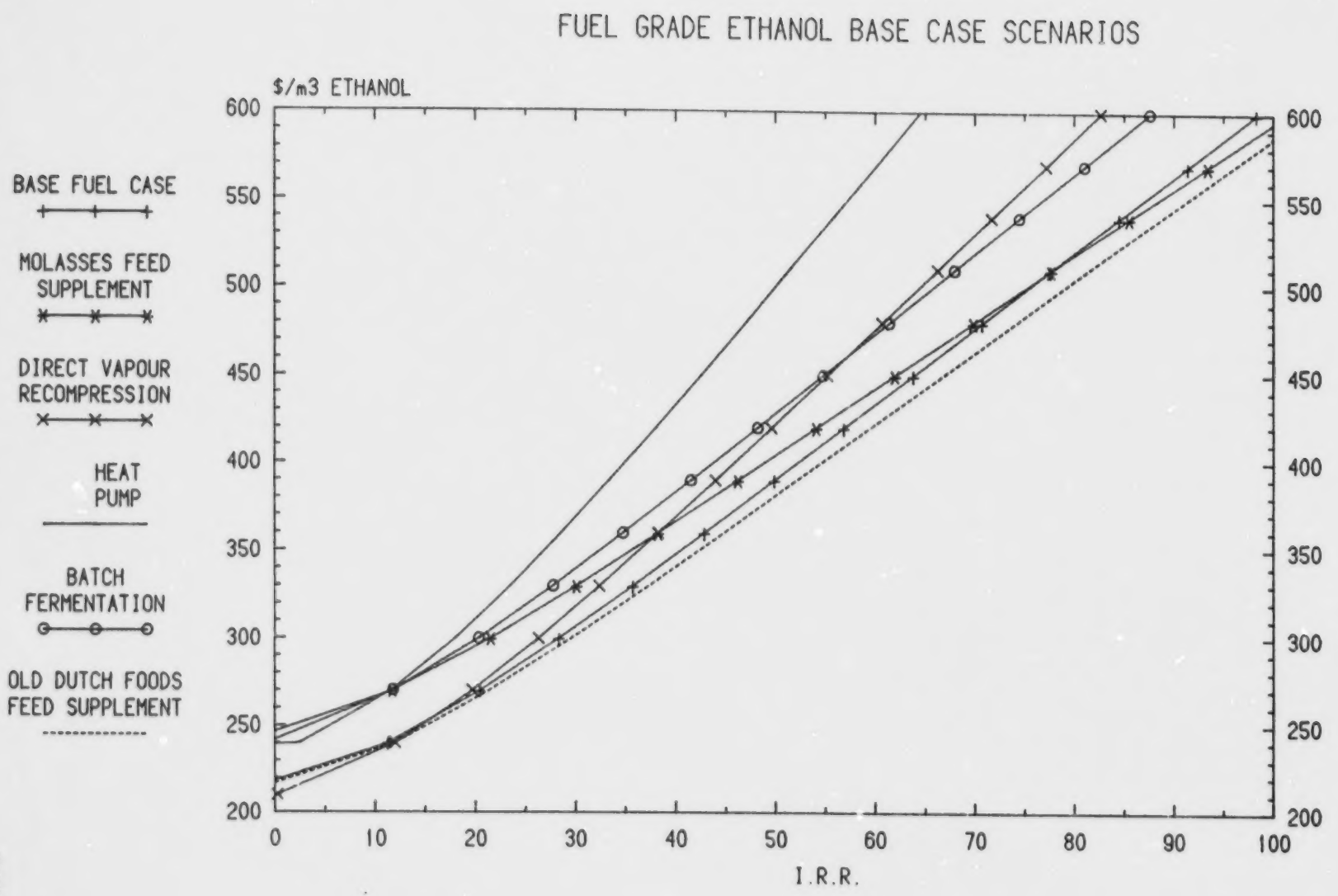
Internal rate of return (IRR) is the value of discount rate that yields a zero present value for the net cashflow. This is calculated by changing the discount rate until zero net present value is obtained.

Scenario and Analysis Results

Several scenarios, for each of the eight fuel ethanol and four food grade ethanol cases, were run to determine sensitivity of the project finances to several parameters. For each scenario, plots of IRR versus ethanol selling price (FOB ethanol plant) were obtained.

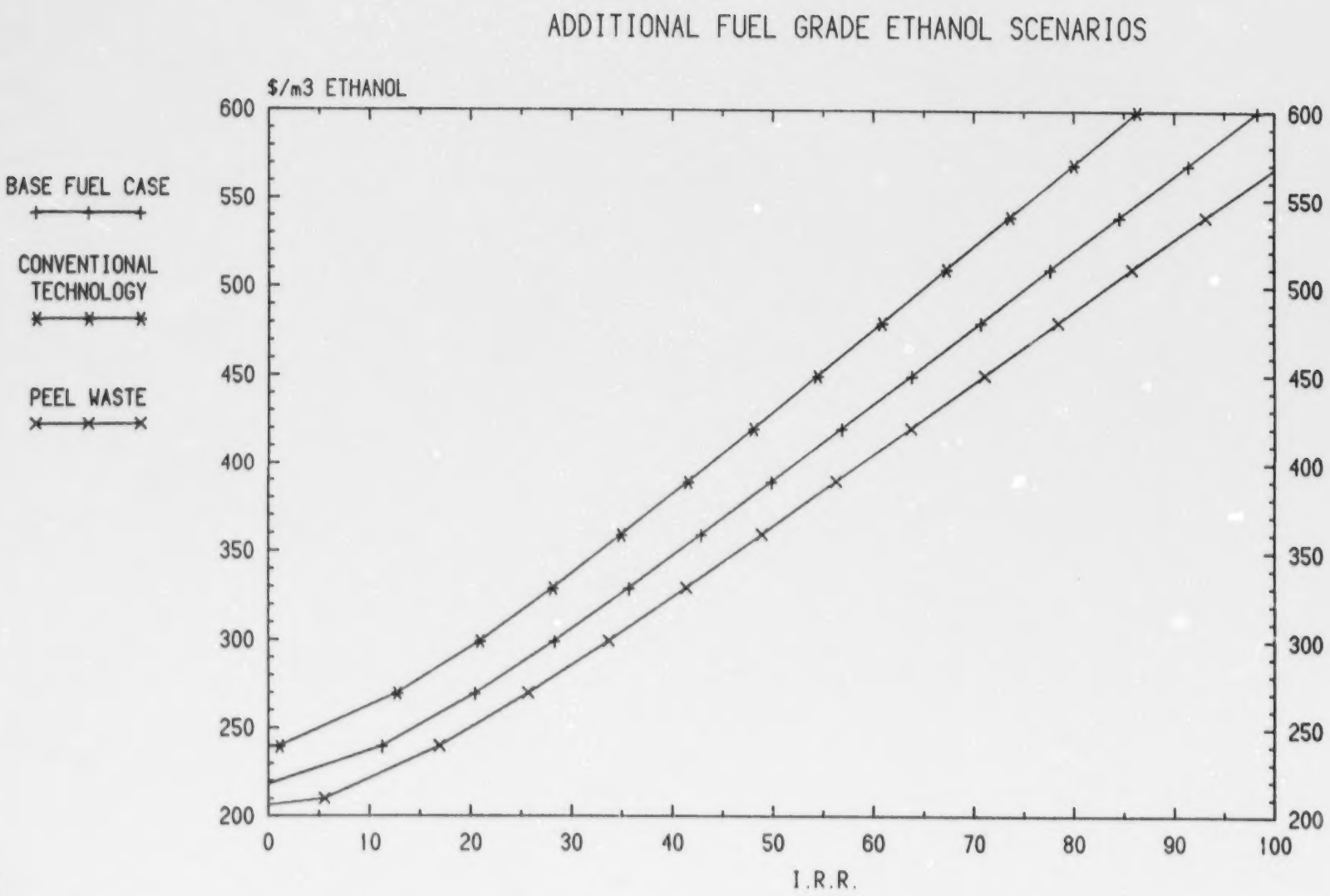
In addition to the cost inputs in Table 7 and 8, the following additional assumptions were made in all cases unless otherwise defined.

- Inflation = 0 percent
- Interest = 4 percent
- 20 year project life
- Capital spent in one year
- Working capital = 25 percent of operating costs, none borrowed
- Debt to equity ratio = 2:1
- Start up cost of $150,000 for fuel grade and $200,000 for food grade scenarios, incurred in Year 2.
- CCA rate of 5 percent for building and site preparation, 15 percent for the rest of the capital costs.



FUEL GRADE ETHANOL BASE CASE SCENARIOS

FIGURE B-1



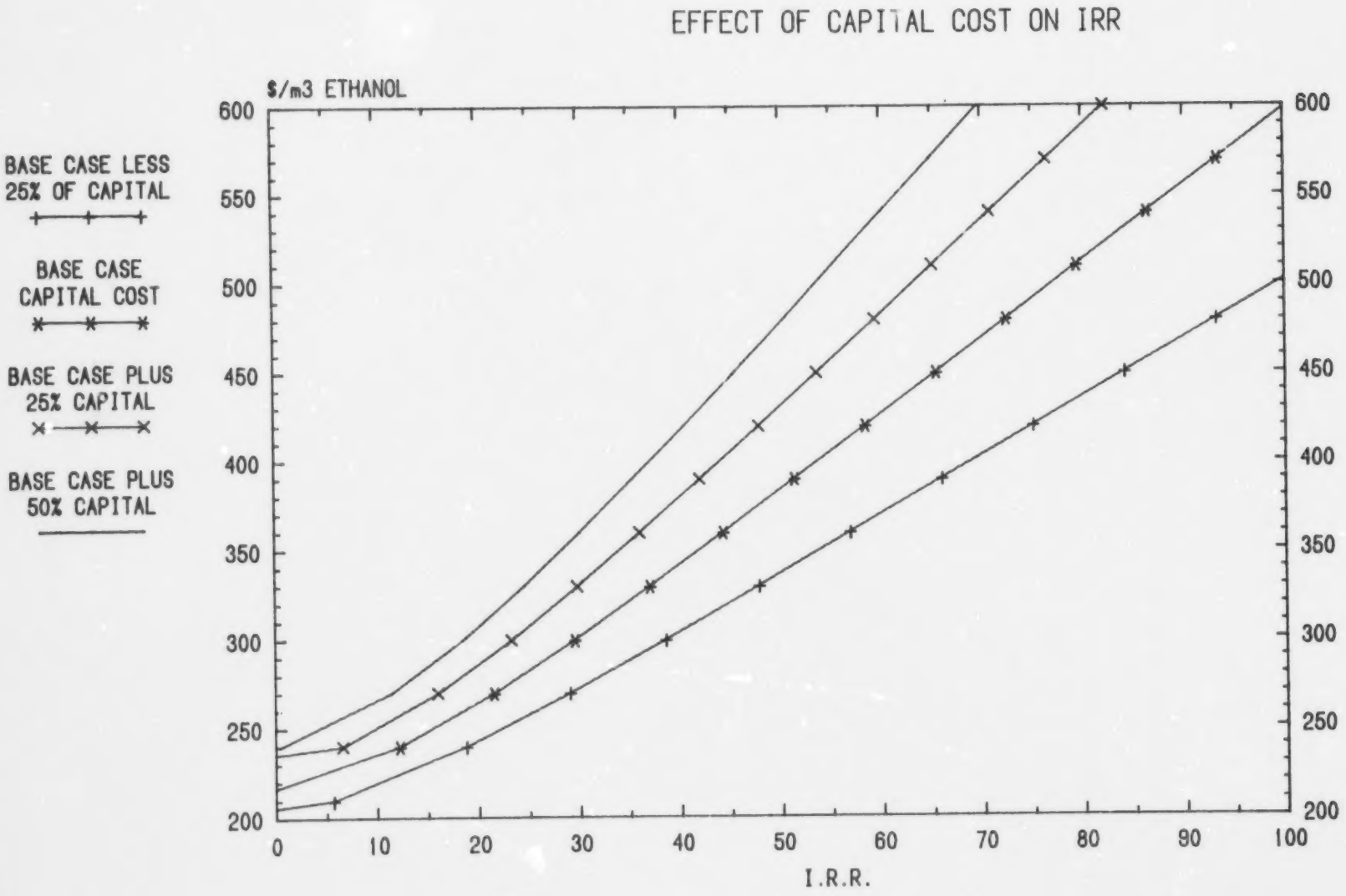
ADDITIONAL FUEL GRADE ETHANOL SCENARIOS

FIGURE B-2

Fuel Ethanol Cases

Each fuel ethanol case previously described was analysed for effect of ethanol price on project IRR. Results of these calculations are shown by Plots B-1 and B-2.

One can see for an ethanol price in the $.30/L range ($300/m^3) that IRR is greatest for the case where Old Dutch Foods' waste starch supplements Carnation's starch feed to the ethanol plant. IRR is slightly less for the base fuel case where Carnation's starch alone is used, that is over 30 percent return. If the estimated of quantities of peel wastes available is increased IRR increases significantly. Where vapour recompression is used to help power ethanol distillation, IRR is slightly less than the base fuel case. IRR is considerably less for any given ethanol price for the other cases, as shown by the graphs.



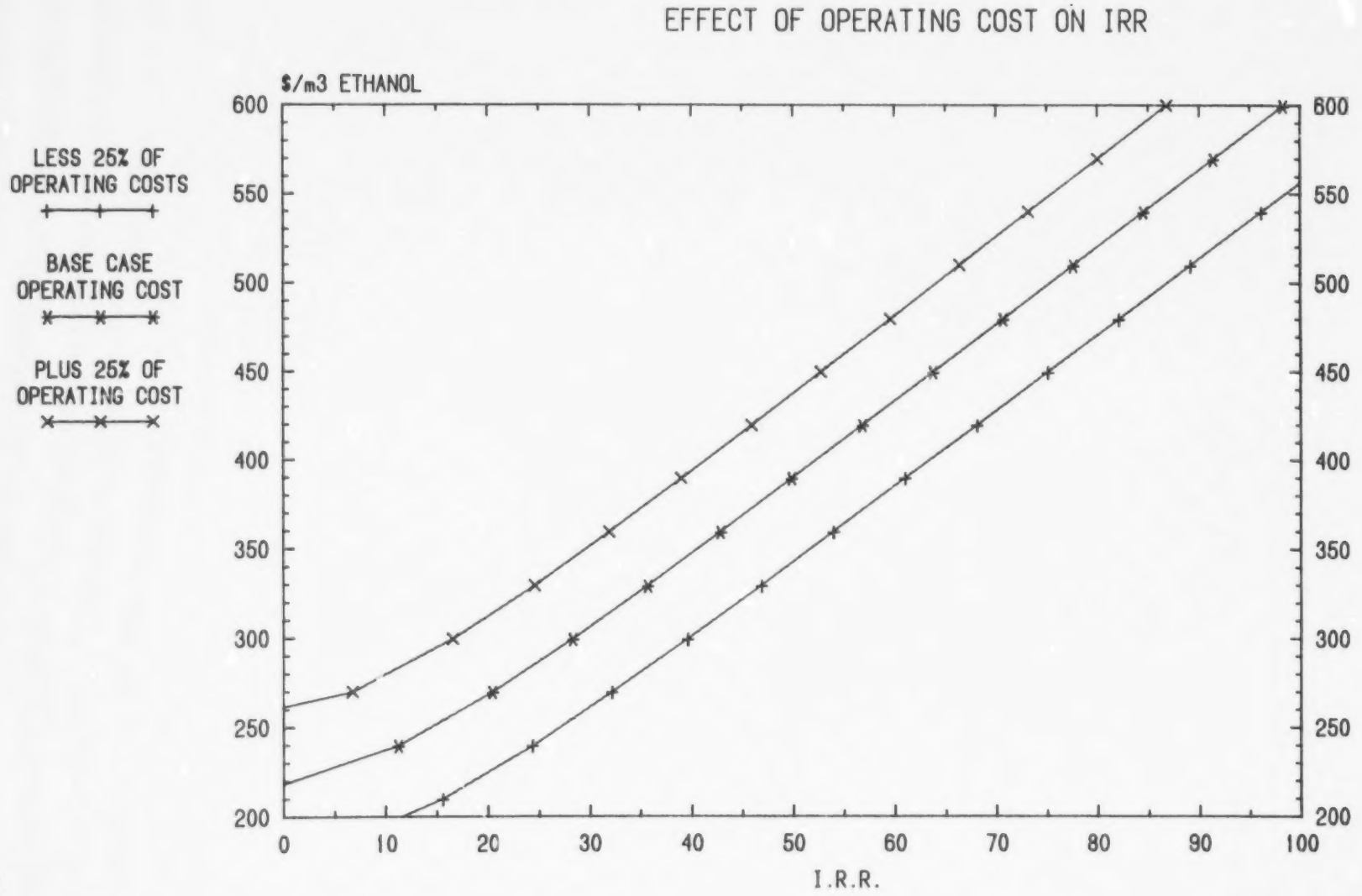
FUEL GRADE ETHANOL BASE CASE SCENARIO

FIGURE B-3

Effect of Capital Cost on IRR

Sensitivity of the base fuel case project finances to capital cost is displayed by Figure B-3. As expected, IRR is very sensitive to changes in capital costs.

Note that capital cost allowance used in this case is 15 percent for all items (total capital cost was entered instead of each separate item to expedite calculations). 15 percent CCA was used for the capital investment depreciation. Thus, a small improvement in IRR is realized due to depreciation of building cost at 15 versus 5 percent annum.

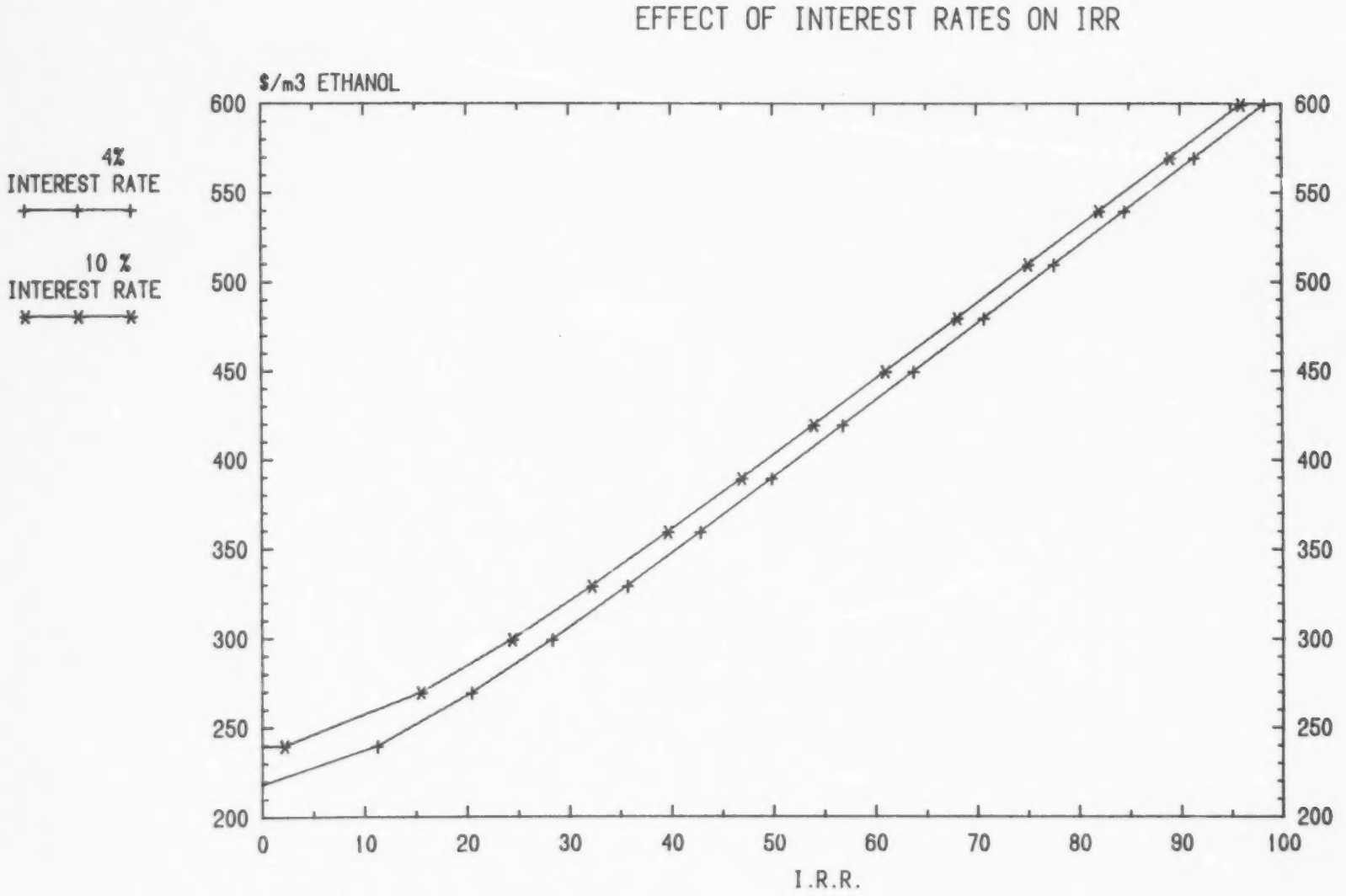


FUEL GRADE ETHANOL BASE CASE SCENARIO

FIGURE B-4

Effect of Operating Cost on IRR

Sensitivity of the base fuel case project finances to changes in operating cost is displayed by Figure B-4. One sees that IRR is highly sensitive to operating costs.

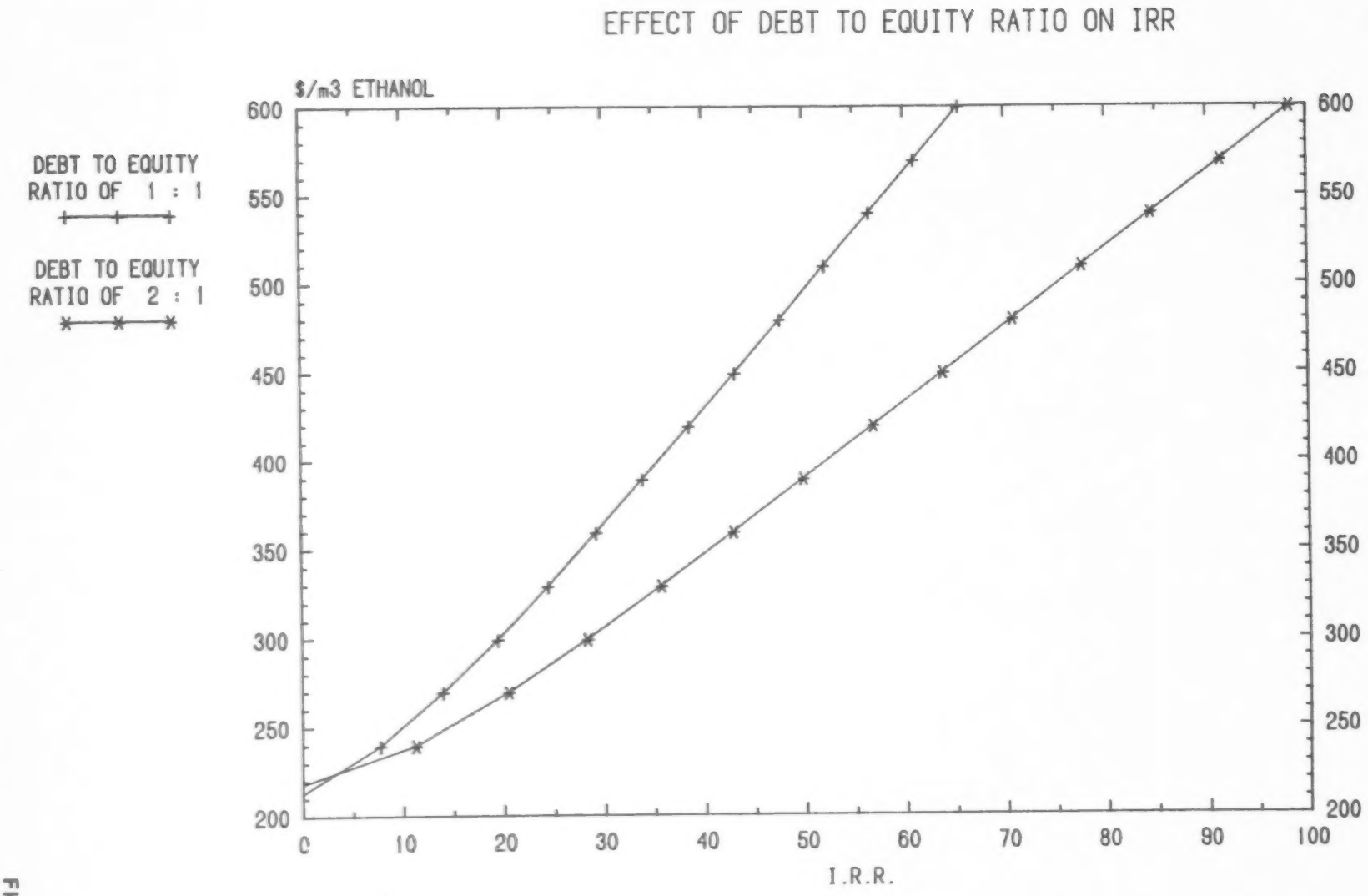


FUEL GRADE ETHANOL BASE CASE SCENARIO

FIGURE B-5

Effect of Interest Rates on IRR

Scenarios where interest rates of 4 percent and 10 percent (0 percent inflation) are calculated as shown by Figure B-5. Project finances are not very sensitive to interest rates despite the large difference (4 versus 10) in interest rates for each scenario.

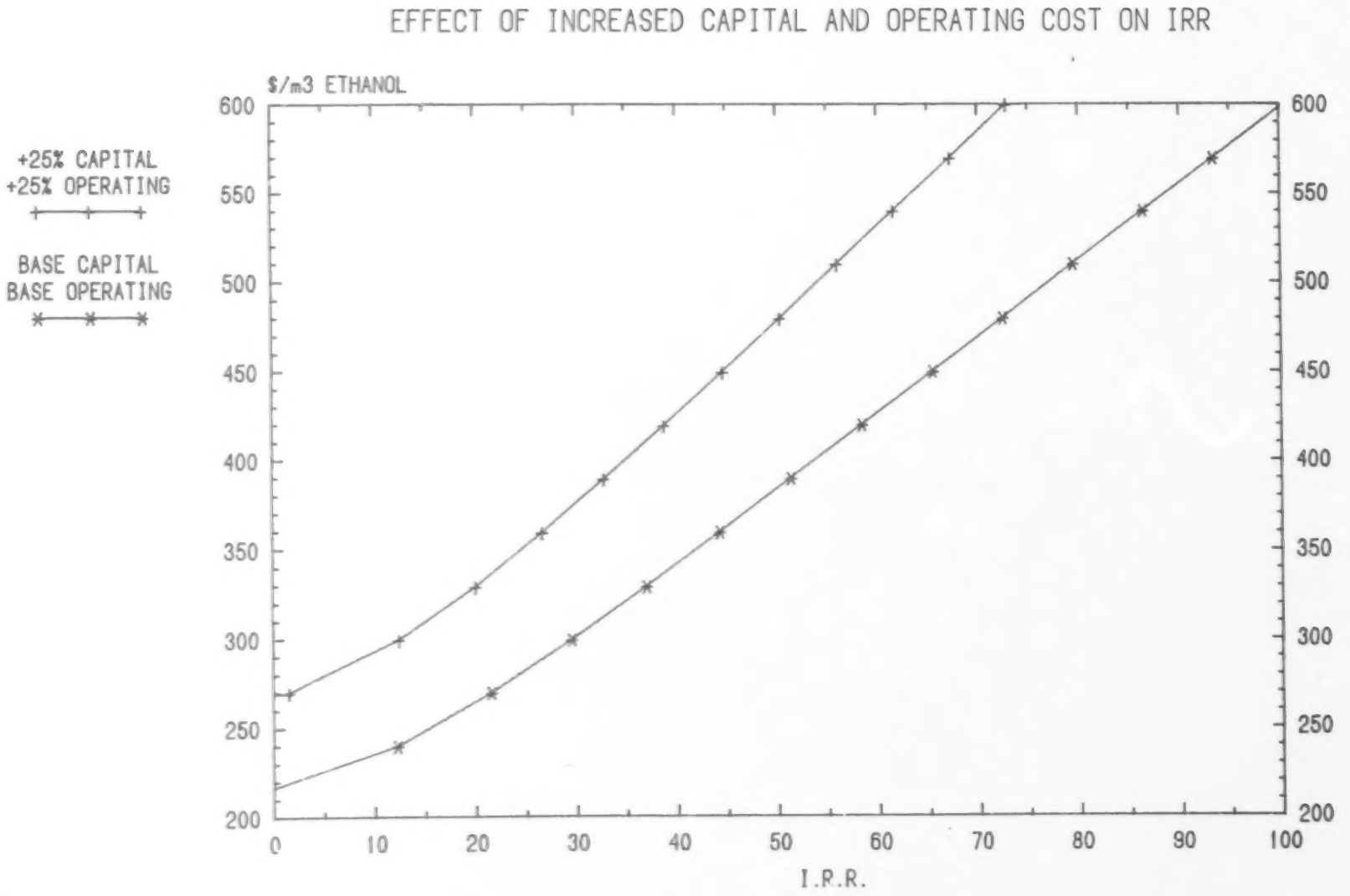


FUEL GRADE ETHANOL BASE CASE SCENARIO

FIGURE B-6

Effect of Debt to Equity Ratio on IRR

Sensitivity of project finances to debt to equity ratio, shown by Figure B-6 is high as one may expect. Using 50 percent equity versus 33 percent equity reduces IRR from about 27 percent to about 19 percent for the base fuel grade ethanol case assuming $.30/L ethanol selling price.

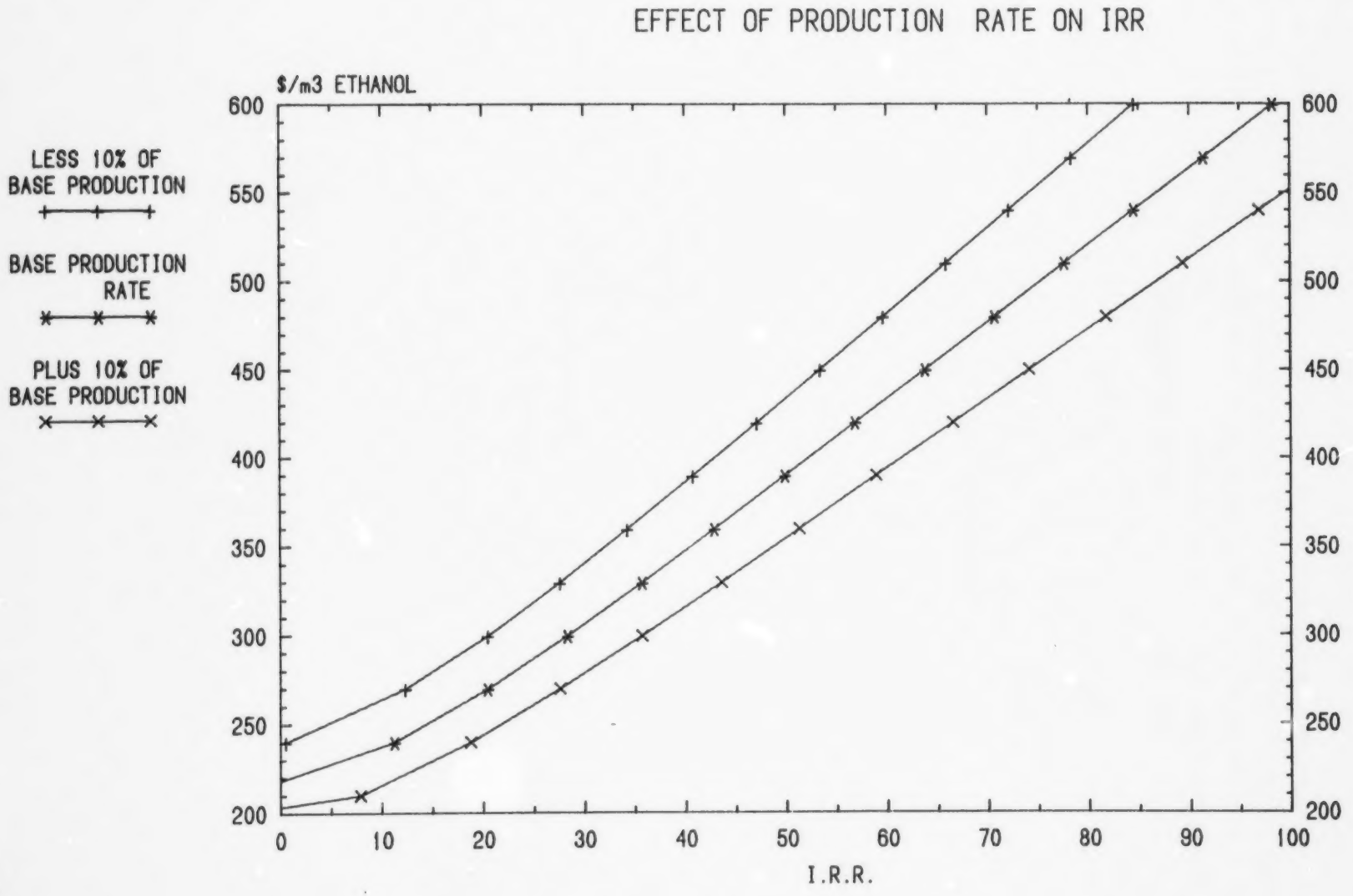


FUEL GRADE ETHANOL BASE CASE SCENARIO

FIGURE B-7

Effect of Increased Capital and Operating Cost on IRR

Figure B-7 shows how IRR can be reduced to as low as about 12 percent for an ethanol selling price of $.30/L if operating and capital costs are both increased by 25 percent over the base fuel case capital and operating costs. This compares to 28 percent IRR if capital and operating costs realized equal those predicted as described previously in this report.

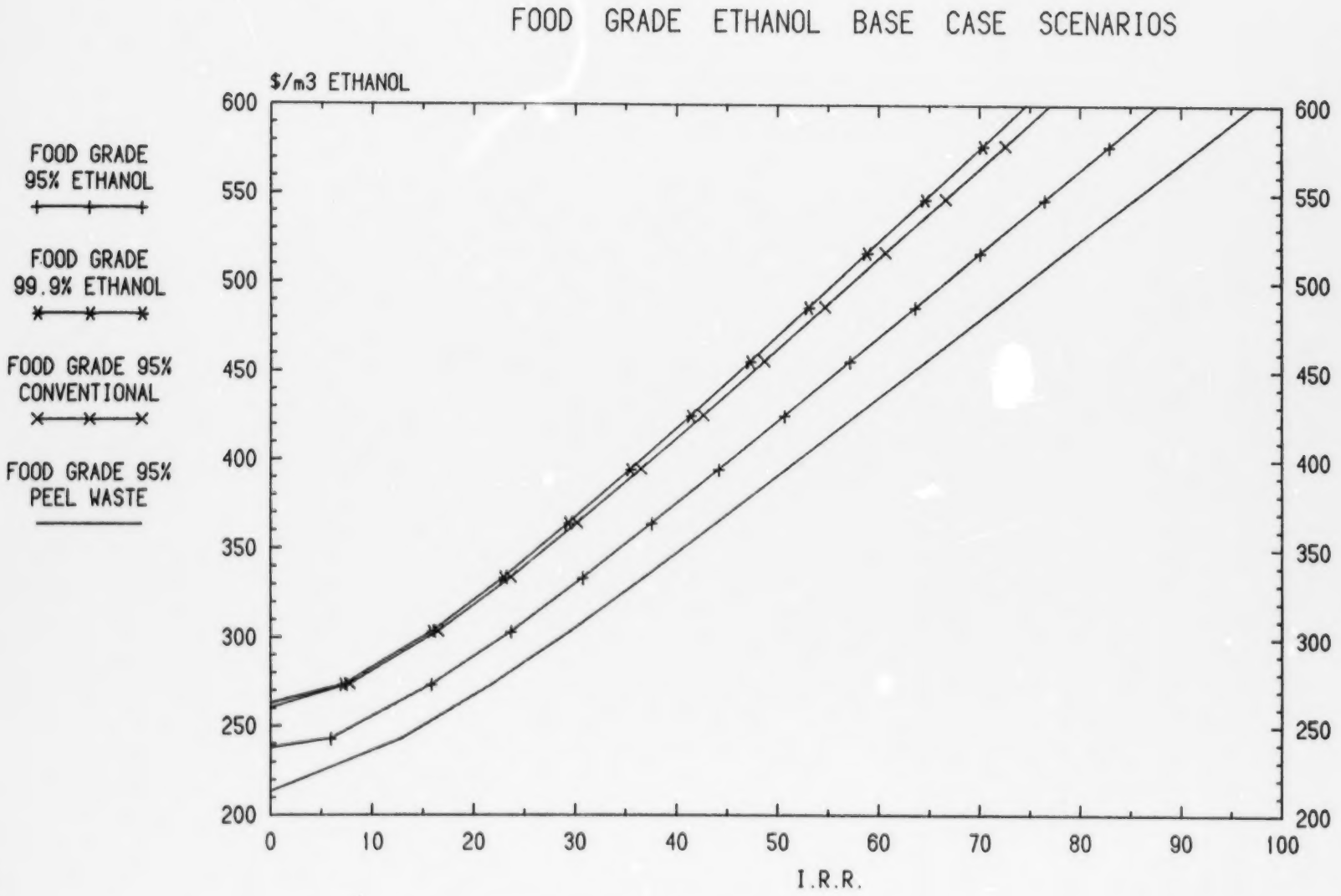


FUEL GRADE ETHANOL BASE CASE SCENARIO

FIGURE B-8

Effect of Production Rate on IRR

Ethanol production rate at the proposed ethanol plant is directly proportional to waste starch production rates at the Carberry site. Effect of ethanol production rates on project IRR is shown by Figure B-8. If production rates should drop to 90 percent of the design rate, IRR falls from about 28 percent to about 20 percent. Conversely, increased starch production with corresponding increase in ethanol production can increases IRR to about 35 percent as shown by the graph.



FOOD GRADE ETHANOL BASE CASE SCENARIOS

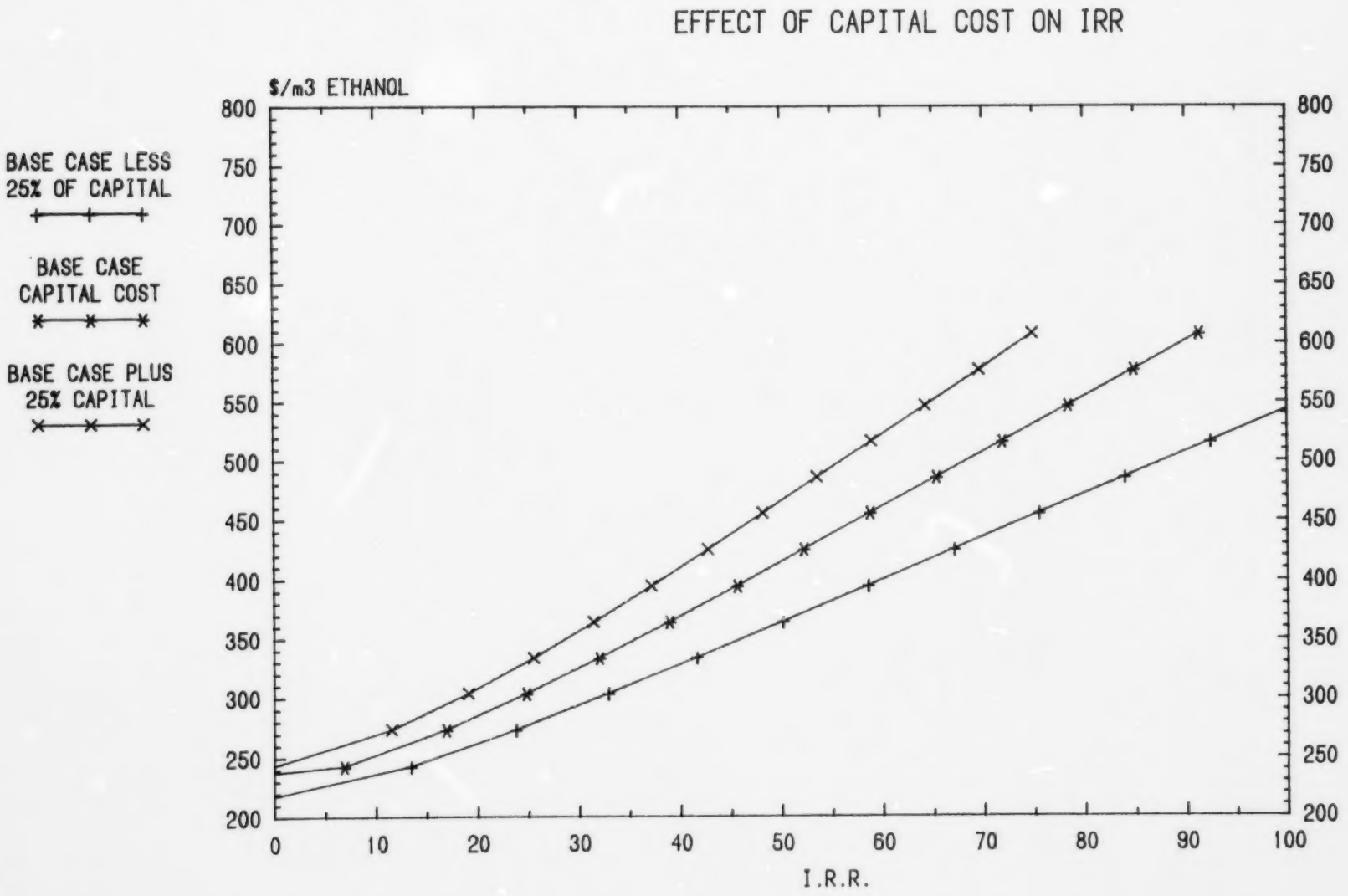
FIGURE B-9

Industrial/Food Grade (95 Percent) Ethanol Case

Displayed in Figure B-9, one sees that furtner dehydrating food grade ethanol from the azeotropic (95 percent) ethanol to anhydrous (99.9 percent), ethanol reduces project IRR. IRR for 95 percent food grade ethanol production with an ethanol selling price of $.35/L (FOB plant) is calculated to be about 34 percent versus anhydrous case of about 26 percent IRR for $.35/L (95 percent) selling price.

If peel wastes quantities are increased, IRR increases to about 40 percent.

Note that for production of 95 percent food grade ethanol, IRR would be about 23 percent for a selling price of $.30/L versus fuel grade case of 28 percent IRR with same ($.30/L) ethanol selling price.



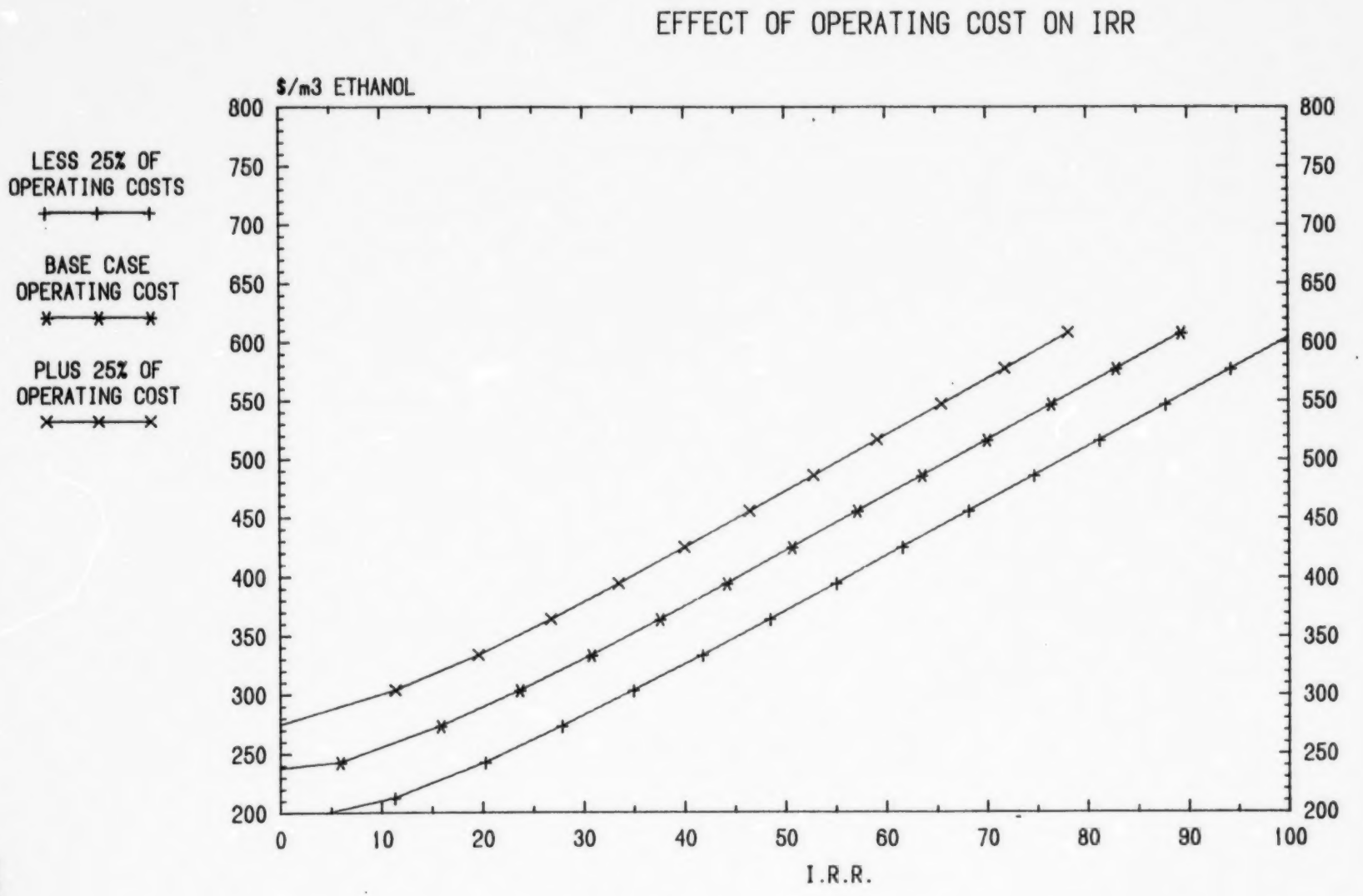
FOOD GRADE ETHANOL (95%) BASE CASE SCENARIO

FIGURE B-10

Effect of Capital Cost on IRR

Effect of capital cost on project IRR for production of 95 percent food grade ethanol is significant as one may expect. Capital cost increase of 25 percent reduces IRR by about 7 percentage points for a $.35/L ethanol selling price as shown in Figure B-10.

Note that capital cost allowances used in this case is 15 percent for all items (total capital cost was entered instead of each separate item to expedite calculations). 15 percent CCA was used for the capital investment depreciation. Thus, a small improvement in IRR is realized due to depreciation of building cost at 15 versus 5 percent annum.

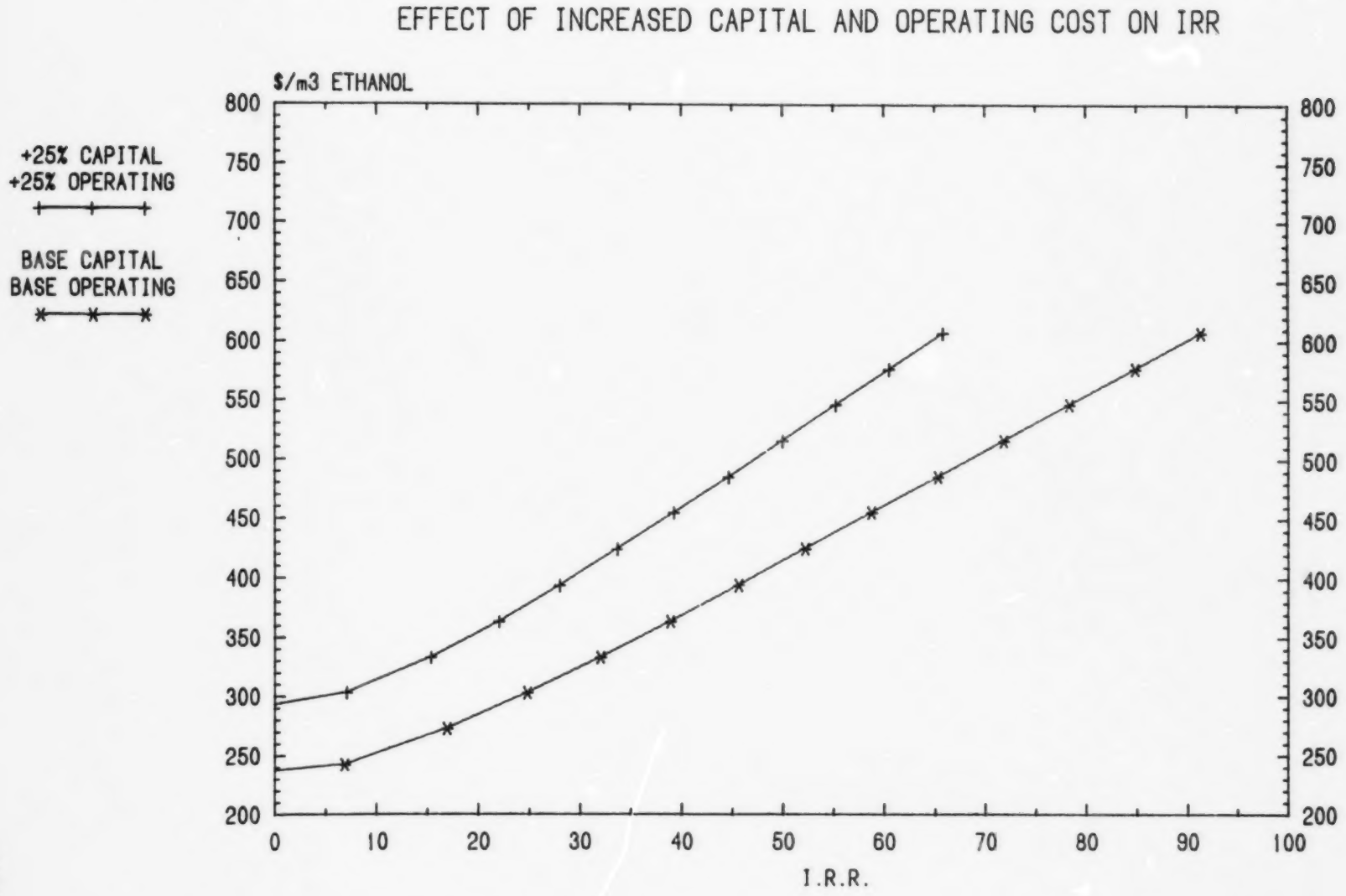


FOOD GRADE ETHANOL (95%) BASE CASE SCENARIO

FIGURE B-11

Effect of Operating Cost on IRR

Effect of operating costs on IRR for 95 percent food grade ethanol production is shown in Figure B-11. IRR ranges from about 46 percent IRR to 34 percent IRR to 23 percent IRR for 75 percent, 100 percent and 125 percent base case capital cost respectively, given a $.35/L (95 percent) ethanol (FOB plant) selling price.

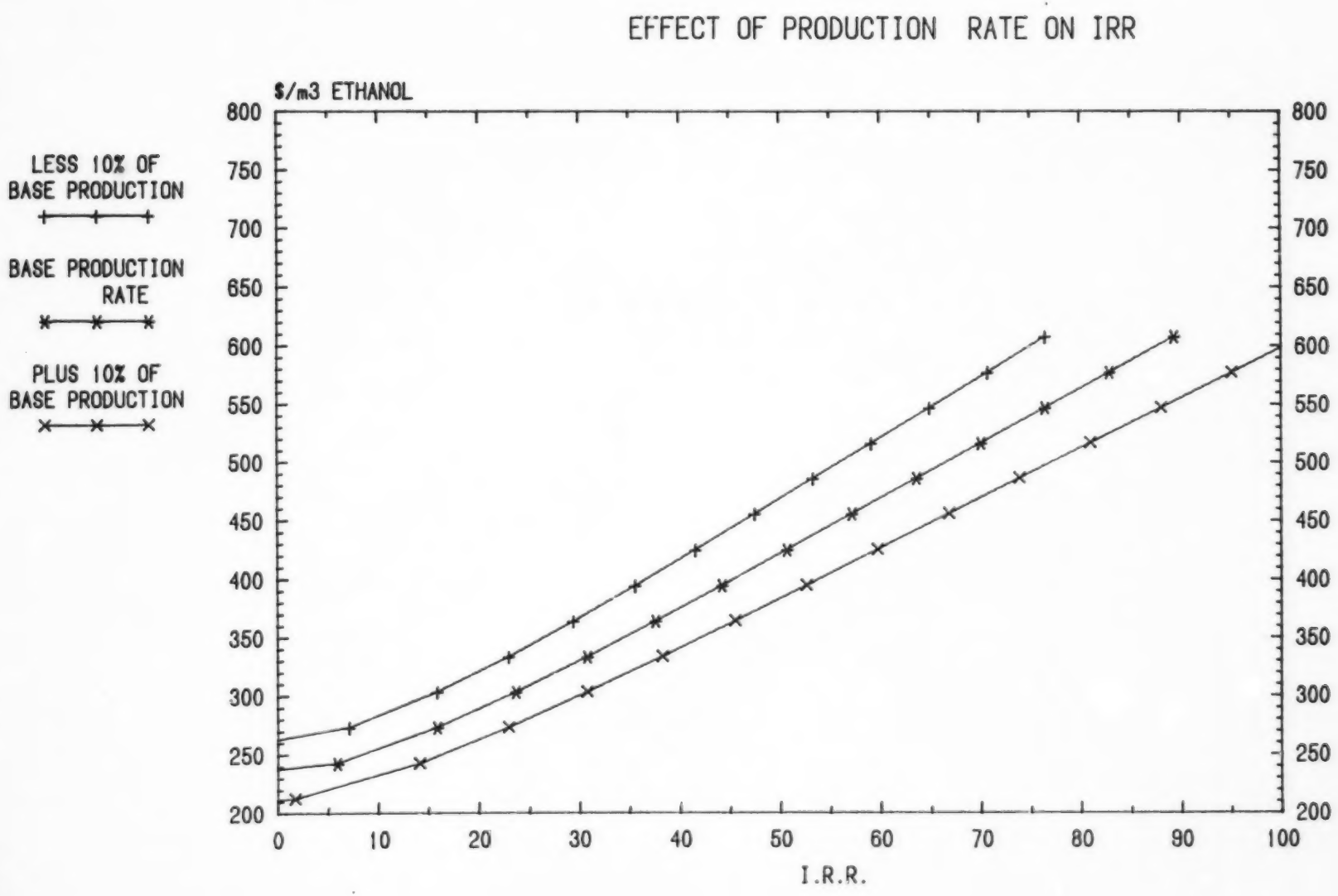


FOOD GRADE ETHANOL (95%) BASE CASE SCENARIO

FIGURE B-12

Effect of Increased Capital and Operating Cost on IRR

As shown by Figure B-12, increase in both operating and capital costs of 25 percent over base case food grade (95 percent) ethanol production decreases project IRR from about 35 percent to about 18 percent for $.35/L (95 percent) ethanol out of gate (FOB plant) selling price.



FOOD GRADE ETHANOL (95%) BASE CASE SCENARIO

FIGURE B-13

Effect of Production Rate on IRR

Effect of ethanol (and thus starch) production rates on project IRR is shown by Figure B-13. For ethanol (and/or starch), production rates of 90 percent, 100 percent and 110 percent of design basis, and for $.35/L (95 percent) ethanol out of gate selling price, IRR is calculated to be about 26 percent, 34 percent, and 42 percent respectively.

Discussion

Based on FOB plant selling prices of 30c/L for anhydrous fuel grade ethanol and $.35/L for food grade (95 percent) ethanol, the production of food grade (95 percent) ethanol appears as having the highest IRR; about 34 percent IRR for the former case, 28 percent IRR for the base case fuel ethanol production case.

Use of ethanol plant starch feed supplement from Old Dutch Foods in Winnipeg enhances ethanol plant production capacity but only marginally increasing rate of return for the project. All other cases yield a lower rate of return than those just described.

Effect of capital and operating costs on project IRR, regardless of which production case, is significant. This highlights the need for accurate cost projections for use in IRR calculations.

Effect of ethanol production rate, and hence starch production rate, on project IRR is also significant. Thus, accuracy in prediction of starch availability and scheduling of same to match ethanol plant production capacity as closely as possible is of significant importance. Increase in availability of starch feed to the ethanol plant can considerably improve project economics.

In summary, for ethanol prices in the $.30/L to $.35/L range, project economies appear favourable, with annual returns in the range of 28 to 35 percent.